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Experimental Study of High Levels of SO2 Removal in Atmospheric-Pressure Fluidized-Bed Combustors

FluiDyne Engineering Corp., Minneapolis, MN

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EXPERIMENTAL STUDY OF HIGH LEVELS OF SO_2 REMOVAL IN ATMOSPHERIC-PRESSURE FLUIDIZED-BED COMBUSTORS

by

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16. ABSTRACT
The report describes tests (conducted in an atmospheric-pressure fluidized-bed combustor (FBC) with a cross-section of 1 x 1.6 m) to demonstrate high levels of SO2 removal when burning a high-sulfur coal and feeding limestone sorbent for SO2 removal. The goal was to achieve 90-plus % SO2 removals with reasonable sorbent feed rates, through suitable reductions in sorbent particle size (to improve reaction kinetics) and increases in gas residence time (to increase gas/sorbent contact time), in a manner predicted by an existing mathematical model. At particle sizes averaging 800 and 1300 micrometers (mass mean), and with gas residence times of 0.5 to 1.5 sec, the measured SO2 retention levels ranged from 88 to 98% when sorbent was fed at Ca/S molar ratios between 2 and 3. This result supports model predictions. Reducing sorbent particle size and increasing gas residence time resulted in modest increases in SO2 removal over the range of conditions tested here. Increases in flue gas O2 content also increased removals. Only one of the three sorbents considered for this project had the attrition resistance necessary to permit use in this testing, indicating that some sorbents will not be suitable for use in dense-phase FBC. NOx emissions ranged from 130 to 236 ng/J during these tests.

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a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Limestone	Pollution Control	13B	08G
Fluidized Bed Processing	Mathematical Models	Stationary Sources		12A
Sulfur Dioxide	Nitrogen Oxide	Fluid-bed Combustion	13H	
Coal	Particles	Particulate	07B	14G
Flue Gases	Fabrics	Fabric Filters	21D	11E
Sorbents	Filtration		21B	07D
	Cyclone Separators		11G	07A, 13I

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ABSTRACT

Tests were conducted in an atmospheric-pressure fluidized-bed combustor having a cross-section of 1 m x 1.6 m (40 in. x 64 in.), for the purpose of demonstrating high levels of sulfur dioxide removal when burning a high-sulfur coal and feeding limestone sorbent for SO_2 removal. The testing was intended to achieve SO_2 removals of 88 percent and greater with reasonable sorbent feed rates, through suitable reductions in sorbent particle size (to improve reaction kinetics) and increases in gas residence time in the bed (to increase gas/sorbent contact time).

The statistically-designed test matrix addressed: one limestone (Greer limestone); sorbent feed rates aimed at two different SO_2 removal rates (88 and 93 percent); two limestone particle size distributions (800 and 1300 micron mass mean); one bed temperature (844 °C, or 1550 °F); two bed depths (0.9 and 1.4 m, or 3 and 4.5 ft); and three gas velocities (0.9 - 1.8 m/sec, or 3 - 6 ft/sec). These conditions resulted in nominal gas residence times in the bed ranging from 0.5 to 1.5 sec. The combustor was nominally operated at 30 percent excess air, although unavoidable variations in excess air levels demonstrated that the excess air level could influence SO_2 retention. Following an initial study of the effects of carry-over recycle and of the number of coal/sorbent feed points, tests were conducted with recycle and with two coal/sorbent feed points.

The measured SO_2 retention levels ranged between 88 and 98 percent when sorbent was fed at calcium-to-sulfur molar ratios between 2 and 3. Retention increased somewhat with decreasing sorbent particle size and increasing gas residence time, and with increasing flue gas O_2 content, over the range tested here. Efforts to identify a suitable sorbent for testing showed that the attrition resistance of the sorbent can play a key role in determining what dense-phase bed depths can in fact be maintained in a combustor, the necessary freeboard height, and other combustor design/operating requirements.

Emissions of NO_x and particulates were also measured during this testing. NO_x emissions ranged between 130 and 236 ng/Joule (0.3 to 0.55 lb/MM Btu). Particulate emissions following the cyclone but upstream of the baghouse ranged between 9 and 35 g/m^3 (4 and 15 grains/SCF); after the baghouse, at the stack, the particle loading ranged between 0.4 and 22 ng/Joule (0.001 and 0.05 lb/MM Btu).

This testing was conducted during the period October 1978 through July 1980.

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1.0 INTRODUCTION

For fluidized-bed coal combustors to be competitive with conventional coal-fired boilers, the fluidized-bed units will have to be able to provide reductions in SO_2 emissions comparable to the reductions possible with conventional boilers using scrubbers. These comparable reductions must be achieved at a competitive cost.

In June 1979, while this study was underway, EPA promulgated New Source Performance Standards (NSPS) for utility boilers which required up to 90 percent SO_2 removal when burning a high-sulfur coal. NSPS were also being evaluated for industrial boilers; this industrial boiler evaluation addressed, as one alternative control level, reductions up to 90 percent. However, prior experimental testing in fluidized-bed combustion facilities had generally focussed on the sorbent feed rates necessary to meet the earlier NSPS for large steam generators which had been promulgated in 1971 (520 ng SO_2/Joule, or 1.2 lb/MM Btu). This earlier standard generally corresponds to a percentage SO_2 removal of only about 80 to 85 percent with a high-sulfur coal. Little experimental work had been conducted with sorbent feed rates necessary to achieve reductions above 90 percent. Accordingly, there was not a substantive data base in large pilot fluidized beds to confirm how a requirement for 90+ percent reduction might impact the design of fluidized-bed combustors, and their capital and operating costs.

In earlier work under EPA sponsorship, Westinghouse Research and Development Center had developed a mathematical model predicting SO_2 removal in a fluidized bed, based upon sorbent/SO_2 reaction kinetics and upon fluidized-bed design and operating parameters (Reference 6). Using this model, it had been predicted that fluidized-bed combustors should generally be able to achieve high levels of SO_2 removal economically, provided that sorbent reactivity is sufficiently great (e.g., through decreases in sorbent particle size), and provided that gas residence time in the fluidized bed (i.e., gas/solids contact time) is sufficiently great, through suitable increases in bed depth and/or reductions in gas velocity (References 7 and 8). The reduced operating costs resulting from reduced sorbent feed requirements would more than compensate for the increased capital costs associated with the larger, deeper combustors that would be needed to provide the increased residence time.

The purpose of the current study is to demonstrate that high levels of SO_2 removal (greater than 90%) can in fact be routinely achieved in fluidized-bed combustors with reasonable sorbent feed rates, if sorbent particle size and gas residence time in the bed are appropriately adjusted. These objectives are to be met through a statistically-designed test program on a reasonably large experimental fluidized-bed combustor which has the flexibility to operate over the range of gas velocities and bed depths needed for this evaluation.

2.0 SUMMARY AND CONCLUSIONS

Twenty-eight coal combustion tests were conducted in a 1 m x 1.6 m (40 in. x 64 in.) atmospheric-pressure fluidized-bed combustor, capable of burning between 55 and 250 kg coal per hour (120 to 550 lb/hr). The purpose of this test matrix was to demonstrate the effects of sorbent particle size and gas residence time in the bed on the ability of fluidized-bed combustors to achieve high SO_2 removals with a high-sulfur coal at reasonable sorbent feed rates (and at reasonable cost), thus validating an existing mathematical model.

The tests were conducted with Illinois No. 6 coal (3.5 % sulfur) and Greer limestone. Two size distributions of the Greer limestone were used, to test their influence on sorbent reactivity: -8 mesh (mass mean about 800 microns) and 4 x 30 mesh (mass mean about 1300 microns). The test matrix covered a range of superficial gas velocities (0.9 to 1.8 m/sec, or 3 to 6 ft/sec) and bed depths (0.9 to 1.4 m, or 3 to 4.5 ft), thus providing nominal gas residence times ranging between 0.5 and 1.5 sec. All tests were conducted at a bed temperature of 844 °C (1550 °F) and an excess air level of 30%, with the coal and limestone being premixed and injected near the bottom of the bed.

Initially, 12 tests were conducted to evaluate the effect of recycle, and of the number of coal/sorbent feed ports (one vs. two), on the retention of SO_2 in the bed. Recycle and the use of two feed ports each improved the retention by 1 to 2 percentage points.

Sixteen additional tests were then conducted without recycle and using two feed ports, to evaluate the effects of gas residence time and limestone feed size, for a range of calcium-to-sulfur molar feed ratios between 2 and 3 (expected to give SO_2 reductions in the vicinity of 90 percent at the test conditions).

For these tests, sulfur retention levels ranged between 88 and 98%. As expected, the highest retention levels were generally achieved at the highest gas residence times (i.e., at deep beds and low gas velocities), the highest sorbent feed rates, the smaller sorbent particle size, and the highest levels of excess air.

Results of a multiple linear regression analysis show that the calcium-to-sulfur molar feed ratio and the excess air were the dominant variables. The sulfur retention increased by about 6% as the Ca/S feed ratio increased from 2 to 3, and by about 3% as the excess air increased from 30% to 40%. Retention increased by about 2% with the decrease in sorbent particle size, and by about 2% with the increase in gas residence time.

Greer limestone was the only one of three sorbents tested which had the attrition resistance needed to permit this testing. Attrition occurred not only in the bed, but also in the solids feed system. Thus, many limestones/dolomites may not be suitable to maintain a dense-phase bed.

NO_x emissions ranged between 200 and 300 ppm, or 130 to 236 ng/Joule (0.3 to 0.55 lb/MM Btu). The highest NO_x levels were measured at the highest excess air and lowest SO_2 levels.

Particulate mass loadings upstream of the baghouse ranged between 9 and 35 g/m^3 (4 and 15 grains/SCF), with 15 to 25% of the particulate smaller than 10 microns. Particulate mass loadings at the stack (downstream of the baghouse) generally ranged between 0.4 and 22 ng/Joule (0.001 to 0.05 lb/MM Btu). Particulate emissions below the NSPS of 13 ng/Joule (0.03 lb/MM Btu) were generally achieved for baghouse air-to-cloth ratios less than about 1.2 $m^3/min/m^2$ (4 $ACFM/ft^2$).

3.0 DESCRIPTION OF TEST FACILITY

The testing was conducted using a pilot-scale coal-fired fluidized-bed air heater built by FluiDyne to test small industrial boiler applications (Figure 1). The combustor is a "vertical slice" of a full-size unit, having the dimensionally correct height and width. Actual prototype conditions of fluidization, bed area-to-feed port ratio, elutriation, and emissions performance are well simulated by the combustor. The combustor, shown schematically in Figure 2, has a bed cross-sectional area of 1.6 m^2 (18 ft^2). During the testing reported here, the crushed coal and sorbent were premixed and fed near the bottom of the bed.

3.1 Fluidized-Bed Combustor

The combustor shell is carbon steel, lined with a castable refractory to inside dimensions of 1 m x 1.6 m (40 in. x 64 in.). The unit is capable of burning between 55 and 250 kg coal per hour (120 to 550 lb/hr). The roof is formed by suspended stainless steel panels backed by block insulation. Adjacent to the hot side of the roof are two water-cooled panel coils.

Combustion/fluidization air enters at the bottom of the fluidized bed, through a stainless steel distribution grid having numerous nozzles. During ignition, the combustion air is preheated to 760 °C (1400 °F). In-bed temperature control is accomplished using an air-cooled tube bundle.

The tube bundle of the in-bed heat exchanger is located between 20 cm (8 in.) and 109 cm (43 in.) above the distributor grid. The distance from the top of the tube bundle to the flue gas outlet is 2 m (6.5 ft). Coal is injected into the combustor near the bottom of the bed at a location just below the tube bundle.

A refractory-lined cyclone, located at the flue gas outlet, may be used to recycle entrained particles to the bottom of the bed. When recycle is not used, a blank-off plate is inserted in the dipleg.

The in-bed heat exchanger consists of 45 stainless steel tubes, each 7.6 cm (3 in.) in diameter, having a bayonet configuration. The air enters through the inner tube and exits in the annular space. The tubes enter the combustor from both ends and are uniformly spaced (Figure 3). Heat removed by the tube bundle can be varied by regulating the cooling air flow. This load changing capability enables operation at velocities between .61 m/sec (2 ft/sec) and 1.83 m/sec (6 ft/sec), with the excess air maintained at 30%.

The flue gas is cooled to 230-240 °C (450-650 °F) by a water cooled heat exchanger located downstream of the cyclone. The inlet temperatures to the baghouse is controlled to 180 °C (350 °F) by tempering with ambient air. During start-up, a reheat burner maintains the flue gas temperature at the baghouse inlet above the dew point to prevent plugging the bags. A precipitator is located in parallel with the baghouse and may be used for particulate collection, but was not used during this program. An induced draft fan located downstream of the baghouse maintains the combustor freeboard at a slightly negative pressure.

The baghouse fly ash collection and bed ash removal are conveyed to a dumpster for disposal. Either stream may be directed into a separate container for flow rate determination. The bed level is maintained constant during operation by making batch withdrawal from hoppers located beneath the distributor grid.

The fuel and sorbent preparation facility provides crushing, sizing, and drying capabilities. Coal is sized to minus 19 mm and pneumatically transported to a storage silo (Figure 2). Likewise, sorbent is prepared to the desired size and stored in a silo. Fuel and sorbent feed hoppers located near the combustor are supported on load cells to monitor feed rates. During operation, they are automatically refilled from the silos by pneumatic transport lines. Variable-speed drive screw feeders beneath the hoppers meter the feed streams. The fuel and sorbent mix as they exit the feed hoppers and pass through a rotary airlock which provides a seal between the pressurized transport air lines and the feed hoppers. The materials then enter the transport line and are conveyed to the combustor feed ports.

Emission tests were conducted with either one or two feed ports. When two feed ports were used, an upflow divider was used to split the flow. In this approach, the coal and limestone are mixed and uniformly distributed in a vertical pneumatic transport line. The divider is located on top of the vertical line and forms a transition to the individual feed lines. A schematic of the feed port locations is given in Figure 4.

3.2 Flue Gas Sampling and Instrumentation

A flue gas sampling system is located at the flue gas duct downstream of the flue gas heat exchanger. A schematic of the sampling system is shown in Figure 5. The system removes moisture and particulates from the flue gas without changing its composition, then directs the gas to the following analyzers.

Gas	Instrument	Ranges
SO_2	DuPont Model 411 UV	0-300 ppm, 0-3000 ppm
NO_2 (cont.)	DuPont Model 411 UV	0-300 ppm, 0-3000 ppm
NO (batch)	DuPont Model 411 UV	0-300 ppm, 0-3000 ppm
O_2	Beckman Model 742	0-25%, 0-10%, 0-5%
CO_2	Beckman Model 864 NDIR	0-20%
CO	Beckman Model 864 NDIR	0-5%, 0-0.5%

There is a single DuPont ultraviolet SO_2/NO_x instrument, which is usually set to monitor the SO_2 in view of the project objectives. Occasionally, the unit is set to monitor NO_2 and to grab a sample for NO analysis (which the instrument does in a batch fashion). To confirm the SO_2 and NO_x results from the UV instrument, grab samples are taken for wet chemistry analysis (EPA Method 6 for SO_2, Method 7 for NO_x), usually once each run.

For measurement of flue gas particulates, an isokinetic sample was drawn upstream of the baghouse (downstream of the cyclone) and passed through an Andersen high-capacity cascade impactor to determine particulate size distribution (see Figure 19 and Section 5.5). Particulate loading downstream of the baghouse was usually measured once each run using EPA Method 5.

4.0 DISCUSSION OF TEST PROGRAM

4.1 General Description

Fluidized-bed combustion tests were run over a range of gas residence times and sorbent particle sizes, to demonstrate the predicted capability for high levels of SO_2 removal. Superficial gas velocities ranged from 0.9 to 1.8 m/sec (3 to 6 ft/sec), and bed depths from 0.9 to 1.4 m (3 to 4.5 ft), to give a range of nominal gas residence times of 0.5 to 1.5 sec. The as-received sorbent -- from the Greer Limestone Co. of Morgantown, West Virginia -- was screened to obtain two sorbent size distributions (see Appendix M): -8 mesh, having a mass mean size of about 800 microns; and 4 x 30 mesh, having a mass mean size of about 1300 microns. The tests were made burning Illinois #6 coal (3.5 % sulfur), with a flue gas excess oxygen level of 5% (about 30% excess air).

The first twelve tests were designed to assess the advantages and disadvantages of using one vs. two coal/sorbent feed tubes, and of recycling vs. not recycling the cyclone catch. Table 1 gives a summary of this initial test matrix (Tests 1-12). The four possible combinations (2 feed point options x 2 recycle options) were tested at two residence times (0.5 and 0.67 sec), for a total of 8 tests. Four additional tests were run without fly ash recycle (at 0.75 and 1.5 sec residence time). In each test, the Ca/S molar feed ratio was adjusted to arrive at a SO_2 retention level close to the 93% target.

The freeboard gas composition profile was surveyed during the shakedown runs made prior to Tests 1-12. Information on the uniformity of combustion helped in selecting the locations of the feed ports, and also helped confirm the accuracy of feed splitting between two coal/sorbent feed ports. The results of this work are discussed in Section 6.4.

Based on the results from Tests 1-12, the balance of the test program was performed without recycle and with two feed ports. Recycle was found to be of only modest benefit to the sulfur retention (giving a 2 percentage point increase in retention) and the combustion efficiency. Illinois coal has a high volatile content and burns with little carbon loss. The number of feed ports also had only a limited effect on sulfur retention (again, 2%) and combustion efficiency. Two-feed-port operation was selected because of the greater levels of freeboard combustion observed with a single port. The most difficulty with freeboard burning occurred at high velocities with the lower bed depth (Section 6.3).

Following the assessment of the results from Tests 1-12, the test program was continued with a series of 16 tests designed to evaluate the effects of gas residence time and sorbent particle size (Tests 13-28). See Table 2. The test matrix addressed: three superficial gas velocities (0.9, 1.4, and 1.8 m/sec); two bed depths (0.9 and 1.4 m); two sorbent feed rates (Ca/S of 2.0 and 2.75); and two sorbent particle sizes (mass means about 800 and 1300 microns).

The test matrix is summarized in Figure 6. Six test conditions plus two replicates were run for each of the two limestone feed sizes. Gas residence times range between 0.67 and 1.5 sec for the -8 mesh (800 micron mass mean) sorbent, and between 0.5 and 1.5 sec for the 4 x 30 mesh (1300 micron) material. The overlap in residence time was provided to permit separation of the effects of feed size.

For Tests 13-28, Ca/S ratios rather than the SO_2 retention levels were fixed, to simplify performance of the tests. Adjusting the Ca/S ratio to attain a specific SO_2 emission level is difficult because of the time required to achieve steady state after each adjustment in the sorbent feed rate. Calcium-to-sulfur ratios of 2.0 and 2.75 were high enough to assure SO_2 retentions greater than 90% at most test conditions. The order of the tests was chosen to minimize the number of times that changes in bed height were required, since it could take several days to effect an increase in bed height.

Prior to the tests with Greer limestone, attempts were made to use two other sorbents -- a limestone from Illinois (Calcium Carbonate Co.) and a dolomite from Minnesota (Lundeen). The attrition/elutriation rate with those sorbents was greater than the sorbent feed rate, making it impossible to control bed level. That experience indicates the difficulty that can be encountered in finding a sorbent suitable for use in fluidized-bed combustors. A discussion of sorbent selection is provided in Sections 6.1 and 6.2.

Even with Greer limestone, unacceptable amounts of carryover were initially experienced when operating with a deep bed at the higher gas velocities. The combustor was modified to add more freeboard, taking care of this problem. See Section 6.2.3.

4.2 Test Procedure

The experimental combustor was operated for 18-hour periods, including a 3-hour bed ignition period and a 1-hour cool-down. During most 18-hour operating runs, steady state SO_2 emission levels were established rapidly enough to complete one test point. The goal was to have at least 6 hours of steady-state operation on point. However, at times fewer hours at steady state were accepted to enable completion of a test point during an 18-hour run.

Once steady state was achieved for the selected test condition, formal data recording began. The flue gas concentrations of SO_2, O_2, CO_2, and CO were continuously monitored and recorded. The DuPont SO_2/NO_x UV analyzer, which generally was set to monitor SO_2, was switched once or twice during the test to grab a sample for NO/NO_2. The integrity of the UV measurements of SO_2 and NO_x was established by wet chemistry measurements (EPA Methods 6 and 7) once during most tests.

The flue gas duct downstream of the baghouse was isokinetically sampled to determine particulate loading, using EPA Method 5, so that baghouse performance could be evaluated. The flue gas duct upstream of the baghouse was isokinetically sampled with a cascade impactor to determine the size distribution in the 1- to 15-micron range.

Samples of the feed coal, feed limestone, bed material, and fly ash (baghouse catch and, when no recycle, cyclone catch) were taken several times during most tests. Samples taken over a test were then composited, and were analyzed as shown in Table 3. A record was maintained of the quantities of material removed from the bed ("bed overflow"), and collected by the baghouse.

A comprehensive set of data were acquired every two hours for calculation of heat balances and combustion efficiencies.

Detailed results from most of these analyses are presented in the Appendices. The results are discussed in Sections 5 and 6.

5.0 DISCUSSION OF TEST RESULTS

Table 4 presents the actual operating conditions for the steady state portion of each test (as distinguished from the design conditions in Tables 1 and 2). Table 4 also presents some of the key test results, including, e.g., SO_2 emissions, SO_2 retention and combustion efficiency. The test numbers in the table correspond to those assigned in Section 4.0 and in Tables 1 and 2. FluiDyne run numbers are also shown; the FluiDyne numbers were assigned sequentially from the start of the project. A total of 57 18-hour FluiDyne runs were needed to successfully complete the 28 test conditions. Most of the extra runs were caused by problems in identifying suitable sorbents, as discussed in Sections 6.1 and 6.2.

The important parameters -- coal and limestone feed rates, Ca/S ratio, SO_2, and O_2 -- were tabulated over 30-minute time intervals. The numbers for these parameters in Table 4 represent the average of those 30-minute averages. A standard deviation is also given for the Ca/S ratio and the SO_2 levels.

Sulfur retention levels were calculated by comparing the sulfur content of the flue gas to the rate of sulfur input with the coal. The sorbent sulfation level was obtained by using the expression for a sulfur balance:

$$\% \text{ Sulfation} = \frac{SO_2 \text{ Retention } (\%)}{\text{Ca/S Ratio}}$$

The SO_2 concentration and SO_2 retention level were sensitive to the variations in oxygen level from test to test. Regression analysis was used to adjust the data to a 5% excess oxygen baseline, providing a better comparison of the SO_2 results.

Many of the detailed results from this test program are presented in Appendices A through P.

5.1 SO_2 Emissions and Sulfur Retention

SO_2 retention levels with Greer limestone ranged between 88 and 98% with Ca/S ratios of 1.9 to 3.0. Figures 7 and 8 show that the data lie on the flat portion of the retention vs. Ca/S profile. A number of variables are identified in addition to the Ca/S ratio. Tests 1-12 were performed with either one or two feed ports, and with or without fly ash recycle. Residence time was a primary variable in Tests 13-28, but was also varied in Tests 1-12. Also, the limestone feed size was varied, depending on the superficial velocity.

The significance of these variables was determined by multivariable regression analysis and is discussed later in this section. The effect of the flue gas excess oxygen level will be discussed first.

5.1.1 Excess Oxygen

The significance of flue gas excess oxygen was apparent in Test 8. The combustor was operated at 6% excess oxygen for a four-hour period, rather than at the desired level of 5%, because of a shift in the calibration of the oxygen sensor. When the excess oxygen level was lowered back to 5%, the sulfur retention decreased from 92 to 89%. In Tests 1-28, the average excess oxygen level varied from 4.2 to 5.5%. These small long-term shifts in the oxygen level, both above and below the average, occurred in most of the tests. Those shifts averaged over 30-minute time intervals correlated well with the averaged SO_2 emissions once at steady state. The correlation shown in Figure 9 for data from Test 9 is typical. Overall, the response of the SO_2 emission to variations in the oxygen level averaged about 10 PPM for a change of 0.1% in the O_2 level (Table 5). Retention levels were adjusted to a 5% oxygen baseline so that the influence of the other variables could be more accurately assessed. The adjustment to the retention level was usually less than 1%.

5.1.2 Regression Analysis

Multi-variable linear regression analysis was applied to the sulfur retention data because of the number of variables and because of the experimental scatter. Table 6 summarizes these results. The most significant variable was the Ca/S feed ratio. The sulfur retention increased about 6% over a Ca/S range from 2 to 3. The response of the sulfur retention to the other four variables examined by regression was only about 2% in each case. The standard error of estimate for the correlation is ± 2.4, so the effects of the secondary variables fall largely within the experimental scatter.

Figure 10 shows the SO_2 retention data adjusted to a gas residence time of 1.0 sec, and plotted as a function of Ca/S ratio and limestone feed size. Effects of the gas residence time are shown in Figure 11, with the retention levels adjusted to a Ca/S ratio of 2.5 and a limestone feed size of -8 mesh. The combustor operation with regard to fly ash recycle and number of feed ports is indicated in the figures.

The influence of the number of feed ports is shown to be slight; however, there was one test with a single feed port in which the performance was quite poor. In Test 12, substantial freeboard combustion occurred and the SO_2 emissions were correspondingly high. The excess oxygen level was increased to 6.1% to reduce freeboard burning and the Ca/S ratio was increased to 3.76, but the sulfur retention was still only 90%. The coal feed was somewhat finer in Test 12, because of segregation in the silo, which may have caused the higher level of freeboard combustion. Retention levels as low as 70% were seen at the start of several tests with two feed ports when one of the feed ports was plugged.

Considerable experimental scatter remained after the regression analysis. Probable measurement errors in the retention level were less than 1%, so a substantial portion of the standard error of estimate came from other sources. Table 7 gives a residuals analysis which suggests that the scatter resulted from not attaining the true steady state retention level in all tests. Changes in the operating conditions which caused the new steady state level to be approached from the high side resulted in a positive residual. When steady state was approached from the low side, the residuals were negative.

Consider the case where the velocity is increased. The residence time is lower so the SO_2 emission level increases. Because less sulfur is captured by the bed, the average bed sulfation gradually decreases, causing the bed reactivity to increase. A transient towards a higher retention level results until a new sulfur balance is established. The time period required to achieve steady state varies greatly depending on velocity, bed depth, Ca/S ratio, and coal sulfur content. The residuals in Tests 1-12 were mostly within ± 2% because the bed depth and velocity were not changed frequently, whereas in Tests 13-28, operating conditions were changed frequently and the residuals ranged from -4 to +5%.

The four test conditions replicated during Test 13-28 are compared in Table 8. In two of the cases, the agreement is poorer than the standard error of the regression analysis. The reason for the poor agreement is that the retention levels were biased on opposite sides of the steady state level. The bed level was increased in Test 15, so the emission level was biased on the high side. In Test 19 the Ca/S ratio was increased, so the emission level was biased on the low side.

Morgantown Energy Research Center has also performed high-level SO_2 retention studies with Greer limestone (Ref. 1). Their test unit was smaller than the FBC used here, but the operating conditions were similar. The agreement between the two studies was quite good, as shown in Figure 12.

5.1.3 Wet Chemistry SO_2

Wet chemistry SO_2 determinations (EPA Method 6) were obtained during most tests for comparison to the UV analysis. The wet chemistry sample was drawn upstream of the sample conditioning system. Results are shown in Figure 13. Agreement was within ± 20% for most of the analyses.

The analyses which were performed during Tests 1-12 are identified separately from those performed during Tests 13-28 because a different technician was used. The results from the initial tests were biased on the high side while the later tests were biased on the low side. The titration endpoint is difficult, so the different results could be due to different interpretations of where the endpoint was. Detailed information on the wet chemistry analyses is given in Appendix B.

5.2 NO_x Emissions and Sulfur Retention

NO_x measurements were made with a Dupont ultraviolet SO_2/NO_x analyzer. These measurements were infrequent due to the limitation that only one species at a time could be analyzed. To maximize the SO_2 data collection, NO_x measurements were made only at the start or completion of a test. The instrument measures both NO and NO_2; however, only NO_2 can be determined continuously.

The NO must be converted to NO_2 with ozone prior to measurement. This step is performed on a batch sample captured in the sample cell. Since essentially all of the NO_x is in the form of NO, the measurements are based on grab samples. The number of grab samples which could be analyzed was limited because each measurement required about 20 to 30 minutes to read a steady state value.

NO_x emissions measured in all tests are plotted in Figure 14 as a function of the excess oxygen level at the time of the grab sample. At excess oxygen levels less than 6.6% (46% excess air), the NO_x emissions are below the New Source Performance Standard of 260 ng/Joule (0.6 lb NO_x/MMBtu) for coal-fired utility boilers. The emission level at 4.8% excess oxygen (30% excess air) is 170 ng/Joule (0.4 lb NO_x/MMBtu). Emissions from Tests 13-28 average higher and are more scattered than emissions from Tests 1-12. A correlation of NO_x emissions vs. SO_2, shown in Figure 15 appears to explain the differences.

In Tests 13-28, some of the SO_2 emission levels were as low as 60 PPM which resulted in significantly higher NO_x emission level. Other variables, such as velocity and bed depth, did not affect the NO_x emission level. The data in Figure 15 was adjusted to a 5% excess oxygen level using the trend lines in Figure 9.

Exxon (Ref. 2) has investigated the reaction of NO_x and SO_2 in a partially sulfated bed of limestone. They hypothesized that an intermediate in the sulfation reaction, $CaSO_3$, is also an intermediate in the NO_x decomposition reaction. At low SO_2 concentrations, the concentration of the intermediate $CaSO_3$ is low, and consequently the NO_x emissions are higher.

NO_x measurements were made using EPA Method 7 as well as the ultraviolet analyzer. The Method 7 gas samples were taken upstream of the sample conditioning during the same time period batch samples were collected in the UV photometric analyzer. A comparison of the UV and wet chemistry NO_x analyses is given in Appendix C. The UV analyses averaged 6% higher than the Method 7 wet chemistry results in Tests 1-12, and 20% higher in Tests 13-28. A different technician performed the analysis during each test series, but both obtained agreement to within several percent on calibration gases.

5.3 Combustion Efficiency

Combustion efficiencies measured during this program ranged between 94 and 98%. For tests without recycle, these data show an influence of superficial velocity. Combustion efficiency values at a velocity of 1.83 m/sec (6 ft/sec) averaged between 1 and 2% lower than at 0.91 m/sec (3 ft/sec). These data are shown in Figure 16.

The use of recycle improved the combustion efficiency on the order of 1 to 2%, with the largest improvement shown at the highest superficial velocity.

The flue gas combustibles heat loss was negligible for these tests.

5.4 Particulate Emissions

The baghouse used for these tests is a reverse-jet pulse type containing 92.9 m^2 (1000 ft^2) of Nomex cloth (Figure 17). Inlet flue gas temperature is controlled to 177 °C (350 °F) by dilution air. A preheat burner prevents condensation during start-up.

The particulate loading downstream of the baghouse was determined once during most tests using EPA Method 5. The sampling site was on a long 0.61 m (24 in.) diameter duct free of flow disturbances. Twelve traverse points located on two perpendicular diameters were each sampled for 5 minutes. The emissions are plotted as a function of the baghouse air/cloth ratio in Figure 18.

Particulate emissions after the baghouse averaged 2 ng/Joule (0.005 lb/MMBtu) at air/cloth ratios between 1.5 and 2.5, well below the New Source Performance Standard of 13 ng/Joule (0.03 lb/MMBtu). At air/cloth ratios between 3.5 and 5.2, the emissions show much scatter and lie on either side of the standard.

Particulate loadings at the baghouse inlet ranged between 9 and 35 g/m^3 (4 and 15 gr/scf), but averaged about the same at each velocity. An examination of the grain loading effects on the scatter at higher air/cloth ratios showed no statistically significant trends.

The emission level during Tests 7, 8, and 13-15 was very high, 56 to 560 ng/Joule, (0.13 - 1.3 lb/MMBtu) because of a baghouse performance problem. Flue gas bypassed the bags at some flanges which support the bags and cages on the plenum divider. Rivets attaching the flanges apparently worked loose with thermal cycling during start-up and shut-down. All flanges were sealed with Gore-Tex and fastened securely with bolts as a fix. No difference is apparent in the emission levels before and after the repair. However, the first two emission levels measured after the repair (the triangles at air/cloth ratios of 3.6 and 4.9 in Figure 18), may be high because of dust which settled in the duct during the poor performance period. The particulate emissions data are summarized in Appendix E.

5.5 Impactor Studies

The flue duct upstream of the baghouse was sampled with an Andersen high capacity cascade impactor to determine the particulate size distribution in the 1 - 15 micron range entering the baghouse. Despite high grain loadings, reasonable sampling times were obtained with the impactor, which had a capacity of several grams per stage. A schematic of the sampler is shown in Figure 19. An in-stack probe was used to draw an isokinetic sample to the impactor, which was enclosed in a hot box. The sample was drawn from the center of the 0.3 m (12 in.) diameter duct, typically for a one hour period.

Figures 20, 21, and 22 show size distribution data at each of the in-bed gas velocities, and Figure 23 compares the trends. The percentage of the particulate smaller than 10 microns ranges from 15% at 1.83 m/sec (6 ft/sec) to 25 percent at 0.9 m/sec (3 ft/sec). This trend is consistent with expectation, since coarser material will elutriate from the bed at the higher velocities.

The size distributions measured at 1.83 m/sec (6 ft/sec) are less reliable than those at the other velocities. High particulate loadings in this case resulted in as much as 70% of the collection being captured in the sample probe upstream of the impactor. The probe and stage 1 collections were always combined in reporting the results, because of the overall coarseness of the probe material. Since finer material was likely included in the probe deposits in the high-velocity runs, the reported results for the high-velocity runs likely under-estimate the fine fraction. The probe collection averaged about 11% of the total at the lower velocities. In-situ use of the impactor with a straight nozzle would have been preferable, but that was not feasible because of the small duct size. Another reason to suspect the data at 1.83 m/sec (6 ft/sec) is that the particulate mass flow rate calculated from the impactor collection averaged 40% less than baghouse collection rates at that velocity. At the other velocities, the mass flow rate determinations agreed to within 4%.

Outputs were determined from the impactor flow rate using a calibration provided by Andersen. Cut points and stage efficiencies were not confirmed by other techniques. However, the appearance of the material on the various stages was distinctive, which suggests a sharp separation. Also, there was no indication of re-entrainment from the first stage to the finer stages. A summary of the impactor data appears in Appendix F.

6.0 DISCUSSION OF TEST OPERATIONS

6.1 Sorbent Selection

A list of sorbents considered for the test program, including their chemical composition, is given in Table 9. Sorbents were chosen primarily from Minnesota, Iowa and Illinois to limit transportation costs. Three sorbents from the eastern part of the United States were also included because of their proven performance in FBC's. An effort was made to locate limestone and dolomite supplies representative of different geographic regions.

A dried and sized product was a requirement which presented difficulties in locating suppliers. A feed size distribution having a surface area mean size of about 500 microns (for use at 0.91 and 1.37 m/sec, or 3 and 4.5 ft/sec) and about 1000 microns (for use at 1.37 and 1.83 m/sec, or 4.5 and 6 ft/sec) was specified in the test plan. The only supplier who could meet these specifications was the Calcium Carbonate Company. As an alternative, suppliers were located who could provide a dried product sized at either -1/4 inch or -1/8 inch. In many cases, it was possible to top or bottom screen the as-received material at the laboratory in such cases to obtain the desired size distribution. Transportation costs made the yield from the screening operation an important consideration.

The first sorbent selected for the test program was the limestone from the Calcium Carbonate Company. This sorbent was the only processed stone available and had been used successfully in the FluiDyne laboratory previously. It was presumed that this sorbent would also perform acceptably in this program; however, that was not the case. Attrition losses were high and it was not possible to maintain a bed at the desired level.

The other sorbents selected for the test program were chosen with the aid of attrition and reactivity screening tests performed by Westinghouse. The attrition tests were made in a preheated bench-scale fluidized bed operating at a velocity of 30 cm/sec (1 ft/sec). The sample was calcined but not sulfated during the test. Results were expressed as the percent of the -8 + 32 mesh sample attrited to less than 32 mesh in one hour of operation. The chemical reactivity tests were performed with a thermogravimetric balance, which was used to determine the sulfation capacity of each stone. A detailed description of the test procedures may be found in References 4 and 6.

A summary of the Westinghouse screening tests is given in Table 10. The four dolomitic stones, identified as Group 1, were evaluated to select the sorbent to be used in place of the Calcium Carbonate Company limestone. The Lundeen dolomite had the greatest reactivity; however, all the candidates were acceptable in terms of reactivity. The primary difference between the sorbents was in the attrition results. Lundeen dolomite had an attrition rate significantly lower than that of the other candidates, and comparable to the rates obtained for the proven reference sorbents. Another point in favor of Lundeen dolomite was that it had been used successfully during a 500-hour demonstration run conducted by FluiDyne previously. The test program was resumed using Lundeen dolomite, but quite surprisingly, high attrition rates were again a problem. The reason was that the shipments received for the test program were a softer stone than had been used previously during the 500-hour test, apparently due to variability in the quarry deposits.

A renewed search for an acceptable sorbent focused on the proven sorbents and on three regional sorbents, identified as Group 2. The regional sorbents had attrition test results significantly greater than for the proven sorbents. This did not mean that the Group 2 sorbents would necessarily be unacceptable. The attrition test results have to be evaluated on a relative basis; however, the few proven sorbents that Westinghouse had tested happened to have had low attrition values. The uncertainties in the attrition test evaluation made it prudent to select one of the proven sorbents. Greer limestone was chosen, and the test program was successfully completed with this sorbent.

6.2 Attrition and Elutriation Problems

Finding an acceptable sorbent for the test program was very difficult. Calcium Carbonate Company limestone and Lundeen dolomite were found to be unusable because of excessive attrition and elutriation of the bed material. Using these materials, the necessary range of bed depths could not be achieved for the full range of operating velocities. Greer limestone was very attrition-resistant and was used successfully, but difficulty was still encountered when a deep bed was used at the higher velocities. The factors affecting performance are discussed below for each limestone.

6.2.1 Calcium Carbonate Co. Limestone

Five runs were made with Calcium Carbonate Co. limestone in attempting to stabilize a bed. At first, limestone was not fed, so that the freshly-calcined sorbent bed material would more quickly attain a sulfated condition. The attrition rates were very high, and the bed level dropped from 1.14 m (45 in.) to 0.53 m (21 in.) during the course of the five runs. Some limestones which are quite soft when calcined, harden once sulfated, but that was not the situation here. The limestone feed rates needed for SO_2 control were insufficient to balance the elutriation losses even at the lowest velocity of 0.9 m/sec (3 ft/sec).

To diagnose the problem, limestone elutriation rates were determined from the chemical analysis of the material collected in the baghouse. The results are given in Table 11. The total limestone elutriation (elutriated bed material plus elutriated feed sorbent) with a 20 x 60 mesh feed size distribution was about 10 percent higher than the limestone feed rate. The rate of bed level change calculated from these elutriation rates compares well to the bed level change actually observed, as shown in the table. The sorbent in the bed elutriated at a rate equal to about 50 percent of the limestone feed rate, while about 65 percent of the limestone in the feed was elutriated. The elutriation of established bed material was determined from analysis of the baghouse collection when limestone was not fed.

A comparison of the size distribution of the baghouse ash versus the limestone feed, illustrated in Figure 24, shows that approximately 25 percent of the feed was fine enough to be quickly elutriated, while another 30 percent is between 40 and 50 mesh, just one step larger.

A coarser, 6 x 20 mesh feed size with the Calcium Carbonate Co. limestone was tested to determine if the feed elutriation losses would be reduced. The elutriation rates for the 6 x 20 mesh feed, shown in Table 11, were about the same as for the 20 x 60 mesh size. The implication was that substantial attrition occurred either during or shortly after calcination, or perhaps in the sorbent feed system.

An attrition test was performed by Westinghouse to evaluate the poor performance of the Calcium Carbonate Co. stone. The attrition rate was very high compared to that of the other sorbents that had been tested, as shown in Tables 10 and 12 (Ref. 4). They also observed that the limestone was a "popper". The term "popper" implies that the CO_2 pressure during calcination causes stones of this type to rupture, or in some cases actually explode into very fine particles. Argonne National Laboratory has performed TGA reactivity screening tests on a large number of limestones. In the course of those tests, they observed the "popping" phenomena in about 20 percent of the stones tested (Ref. 3).

The high elutriation rates observed in these tests conflict with results FluiDyne obtained from tests conducted previously. The elutriation rates for 20 x 60 mesh Calcium Carbonate Co. limestone were measured to be about 50 percent of the limestone feed rate in one previous case, for tests conducted with no recycle and a velocity of 1.37 m/sec (4.5 ft/sec). The limestone was also used successfully during a long-term corrosion test made at 0.79 m/sec (2.6 ft/sec). The reason for the different attrition performance of that sorbent in this study is not known; however, it is believed that differences in the quarried product are the cause.

6.2.2 Lundeen Dolomite

Performance problems encountered with Lundeen dolomite were very similar to those discussed for Calcium Carbonate Co. limestone. The bed calcination and sulfation steps resulted in the bed level dropping from 1.1 m (43 in.) to 0.47 m (19 in.); and, the elutriation rates could not be balanced with a reasonable dolomite feed rate.

The attrition problem with Lundeen dolomite was very much unexpected because of FluiDyne's past experience with the sorbent, and also because of a favorable Westinghouse attrition evaluation. In the previous FluiDyne 500-hour run, this dolomite was used as-received, sized -6 mm (-1/4 in). About 50 percent of the feed was fine enough to be quickly elutriated, but no difficulty was encountered maintaining stable bed depths at a velocity of 1.2 m/sec (4 ft/sec) in the previous testing.

For the current program, the as-received material was screened to 10 mesh x 60 mesh to obtain a product having a 500 micron surface area mean size. Despite this screened product being considerably coarser than the as-received material used in the earlier testing, the elutriation rates in this study were much greater than observed previously. Along with greater attrition in the bed, pneumatic transport of the dolomite between the screening equipment, storage silo, and feed hopper resulted in substantial attrition. The size distribution at the feed hopper, however, was very similar for these tests and the earlier 500-hour demonstration test (Figure 25).

A variable quarry product is believed to be the reason for the high attrition rates. The dolomite shipment received for these EPA-sponsored tests, when compared with that delivered earlier, appeared much more friable. Westinghouse evaluated both the old and new shipments in an effort to confirm this. They found that the shipments did have compositions differing significantly from each other (Table 13). The relative attrition resistance of the different sorbents tested by Westinghouse is not always consistent with the FBC test experience. All the Lundeen dolomite samples evaluated had attrition rates comparable to the rates for the proven reference sorbents (Table 12). Samples of the new shipment, both dried and undried, were evaluated to confirm that the rotary dryer did not have an affect on the strength of the material. One test was made in an unheated test unit to determine the affect of calcination on the attrition behavior.

In addition to the Westinghouse attrition testing, the sorbents were also subjected to a hardness test referred to as the Los Angeles abrasion test. This test was performed on samples of the old and new Lundeen dolomite shipments, and on Greer limestone. The attrition loss for the old dolomite shipment, according to the Los Angeles test, was midway between the losses for the two other samples (Table 14).

6.2.3 Greer Limestone

Greer limestone is a very attrition-resistant stone; however, high elutriation rates were still encountered with a deep bed when recycle was not used. The difficulty was that the freeboard height of the experimental combustor was too low, leading to carryover of coarse bed particles. The combustor was constructed with a compact freeboard, in view of the space limitations in many industrial facilities. The original freeboard height had been 2.28 m (7.5 ft) with a 1.37 m (4.5 ft) bed depth; however, prior to this EPA-sponsored testing, it was decreased to 1.07 m (3.5 ft) during the course of test work done in support of a demonstration plant design project.

Tests 1-4 -- which were performed with no recycle and with a bed depth of 0.9 m (3 ft) -- had acceptable elutriation rates despite the low freeboard. In Test 5, the bed level was raised to 1.37 m (4.5 ft), with the result that the elutriation rates increased greatly and bed level control was not possible.

To reduce the elutriation rates, the roof was returned to its original height of 2.28 m. This involved raising the stainless steel roof panels and bricking the exposed side-walls. The modification reduced elutriation to the extent that the limestone feed rate and elutriation losses were balanced, allowing the deep bed tests to be successfully completed.

The effect of operating conditions on the elutriation rate is shown in Figure 26 for the 27 tests with Greer limestone. The bed elutriation rates were determined from the fly ash calcium content, and are expressed as a Ca/S feed rate equivalent. The limestone elutriation with a 0.9 m (3 ft) bed depth averaged 40 percent of the limestone feed rate with no recycle, and 33 percent of the feed rate with recycle. Limestone elutriation and feed rates were about equal with a 1.37 m (4.5 ft) bed depth at velocities of 1.37 m/sec and 1.87 m/sec (4.5 ft/sec and 6 ft/sec). Velocity did not have a significant affect on the percentage of limestone feed elutriated because the higher bed attrition rates were offset by the higher limestone feed rates.

Typical size distributions for the elutriated material with no recycle are shown in Figures 27 and 28. The maximum particle size of the fly ash agreed well with terminal velocity predictions at both 0.9 and 1.83 m/sec (3 and 6 ft/sec) when the bed depth was 0.9 m (3 ft). In the case of a 1.37 m (4.5 ft) bed depth, particles as much as four times coarser were able to escape. Several rows of cooling tubes were exposed above the shallower bed, which may have reduced the effect of freeboard splashing.

6.3 Freeboard Combustion

It was apparent from high freeboard gas temperatures that combustion was occurring above the bed in tests made at velocities of 1.37 and 1.83 m/sec (4.5 and 6 ft/sec). The freeboard emission surveys described previously showed that combustion nonuniformities would exist to some extent despite optimal positioning of the feed tubes and accurate feed splitting. Factors which increased the above-bed combustion were the same as those which would increase the combustion nonuniformity.

The strong velocity dependence of above-bed combustion, as determined from the temperature difference between the flue gas and bed, is shown in Figure 29. The flue gas temperature in the freeboard was tempered slightly by two water cooled panel coils mounted beneath the roof. Above-bed combustion was negligible at 0.9 m/sec (3 ft/sec); however, higher velocities and operation with one feed tube increased combustion above the bed.

A 60 $^{\circ}$C (108 $^{\circ}$F) rise in the flue gas temperature, as observed at 1.83 m/sec (6 ft/sec) with two feed tubes, requires that approximately 3 percent of the heat input be released above the bed. The average carbon monoxide concentration required at the top of the bed for the equivalent combustion is approximately 0.7 percent. Combustion occurs rapidly as the bubble and emulsion gases mix after leaving the bed, and therefore the average carbon monoxide concentration at the traverse axis used in the freeboard surveys would be lower than at the bed surface. No carbon monoxide was detected downstream of the flue gas heat exchanger, the sampling location for the flue gas instruments, indicating that combustion in the freeboard was essentially complete.

Above-bed combustion created operating difficulties during the initial Greer limestone runs at the conditions for Test 1. Flue gas temperatures up to 945 °C (1750 °F) were encountered with a 1.37 m/sec (4.5 ft/sec) velocity when the excess air level was 25-30 percent. Temperatures that high endangered the stainless steel construction of the flue ducts, cyclone, and flue gas heat exchanger inlet. The flue gas temperature rise of 114 °C (200 °F) lies well above the range of the data at 1.37 m/sec (4.5 ft/sec) in Figure 29. These exceptionally high temperatures may have resulted from use, at the time, of a divider plate device for feed splitting.

After a sequence of runs with high flue gas temperatures, several modifications were made to the system. A spray nozzle was installed at the inlet of the flue duct to limit the temperature of downstream components. In addition, the upflow feed divider was installed to obtain a better split. The additional water vapor in the flue gas did not affect the emissions readings which are measured on a dry basis.

The strong influence that fuel distribution can have on performance was illustrated in several tests made with two feed tubes, where one tube was initially plugged. With one tube plugged, the SO_2 emission level was two to three times higher than it was with the two tubes operating properly, and the flue gas temperature was 950 °C (1750 °F) or higher. Once the plugged feed tube was blown clear with compressed air, and balanced feed restored, the SO_2 emissions and the flue gas temperature returned to their normal levels.

6.4 Air/Fuel Distributions - Freeboard Flue Gas Surveys

In a large FBC, the number and location of feed tubes can significantly influence performance. Zones of intense combustion at the feed tubes may result in a deficiency of oxygen at those locations and an excess elsewhere. Local reducing zones can adversely affect sulfur retention and may promote corrosion.

During the first 12 tests, either one or two feed tubes were used to feed the 1.65 m^2 (18 ft^2) bed area. To be certain that the comparison on the basis of sulfur retention would be meaningful, performance evaluations were made by surveying the freeboard emissions profile. One objective was to identify favorable positions for the feed tubes. A second objective was to confirm accurate splitting of the feed stream when two feed tubes were used.

6.4.1 Freeboard Emission Survey - Single Feed Tube

Two feed tube locations were evaluated for use in the single feed tube tests. One location was near the wall and the other extended into the center of the bed. The vertical position of the feed tubes was 0.15 m (6 in) above the fluidizing grid, just below the in-bed heat exchanger tubes.

Figure 30 shows oxygen profiles obtained 0.71 m (2.3 ft) high in the freeboard. Combustion was much more uniform with the feed tube located in the center of the bed. The feed location near the wall resulted in a range in freeboard oxygen concentrations of 1% to 14%.

Oxygen profiles obtained during operation at 0.9 and 1.37 m/sec (3 and 4.5 ft/sec) with a feed tube in the center of the bed are shown in Figure 31. The effect of a higher velocity and, consequently, a higher load was a slight increase in the oxygen nonuniformity. An emissions profile which includes CO and SO_2 as well as oxygen is shown in Figure 32. The lump in the CO emissions identifies the region adjacent to the feed tube where the volatiles are released and rapidly burned. SO_2 emissions occurred solely in the half of the bed opposite the feed tube. The flatness of the oxygen profile opposite of the feed tube differs somewhat from the other profiles, which have more of a bell shape. However, the minimum oxygen level in all cases was in the vicinity of the feed tube.

6.4.2 Freeboard Emissions Survey - Two Feed Tubes

Several methods of feed splitting were evaluated for use in the two feed port tests. At first, the split was made with a divider plate located in the drop-leg connecting the airlock and pneumatic pickup points (see Figure 33). A freeboard emission profile obtained with this system is shown in Figure 34.

The nonuniformities which are evident were caused by an unequal coal/limestone distribution at the splitter. The side of the divider nearest the coal hopper received an unequal share of coal, while the other side received an unequal share of limestone. The splitter design was modified to include adjustable baffle plates for directing the solids flow onto the divider. An accurate coal split could be obtained by adjusting the baffles to give a uniform oxygen profile in the freeboard; however, several limitations were evident. The split was dependent on load, making adjustments necessary in the baffle positioning as load changed. In addition, the limestone split could not be determined.

An "upflow" feed splitter was installed to overcome these problems. The upflow splitter is shown schematically in Figure 35. The design is similar to the splitter used by Babcock & Wilcox (Ref. 5) on the 1.8 m x 1.8 m (6 ft x 6 ft) combustor. The coal and limestone are mixed and uniformly distributed in a vertical pneumatic transport line. A divider at the top of the vertical line forms a transition to the individual feed lines. After installation, the system was checked by collecting the material fed in separate barrels. The deviation in the feed rates was less than 4% from the average (Table 15). A freeboard oxygen profile obtained with the upflow splitter is shown in Figure 36. The region between the feed ports had an excess of oxygen, but overall, combustion was reasonably uniform.

Freeboard oxygen profiles acquired during operation with one and two feed tubes, after installation of the upflow splitter, are compared in Figure 37. There was only a slightly greater non-uniformity using two feed tubes than for a single feed tube.

7.0 REFERENCES

1. Mei, J. S., et al., "Fluidized-Bed Combustion of High-Sulfur Coal," Morgantown Energy Research Center, U. S. Department of Energy, METC/RI-79/4, April 1979.

2. Skopp, A., et al., "Studies of the Fluidized Lime Bed Coal Combustion/Desulfurization System," report prepared for the U. S. Environmental Protection Agency by Esso Research and Engineering Co., APTD-1116, December 1971 (NTIS No. PB 210-246).

3. Johnson, I., et al., "Support Studies in Fluidized-Bed Combustion: 1978 Annual Report," report prepared for the U. S. Environmental Protection Agency and the U. S. Department of Energy by Argonne National Laboratory, EPA-600/7-79-203, August 1979 (NTIS No. PB 80-112 758).

4. Newby, R. A., Westinghouse Research and Development Center, personal communication, April 27, 1979.

5. Bonk, D. L., "Stream Splitter for Pneumatic Transport of Crushed Solids," presented at the International Powder and Bulk Solids Handling Processing Conference, Rosemont, IL, May 1978.

6. Keairns, D. L., et al., "Fluidized-Bed Combustion Process Evaluation: Phase II - Pressurized Fluidized-Bed Coal Combustion Development," report prepared for the U. S. Environmental Protection Agency by Westinghouse Research and Development Center, EPA-650/2-75-027c, September 1975 (NTIS No. PB 246-116/AS). Appendix E.

7. Newby, R. A., et al., "Effect of SO_2 Emission Requirements on Fluidized-Bed Combustion Systems: Preliminary Technical/Economic Assessment," report prepared for the U. S. Environmental Protection Agency by Westinghouse Research and Development Center, EPA-600/7-78-163, August 1978 (NTIS No. PB 286-971).

8. Ahmed, M. M., D. L. Keairns and R. A. Newby, "Effect of Emission Control Requirements on Fluidized-Bed Boilers for Industrial Applications: Preliminary Technical/Economic Assessment," report prepared for the U. S. Environmental Protection Agency by Westinghouse Research and Development Center, EPA-600/7-81-149, September 1981 (NTIS No. PB 82-221763).

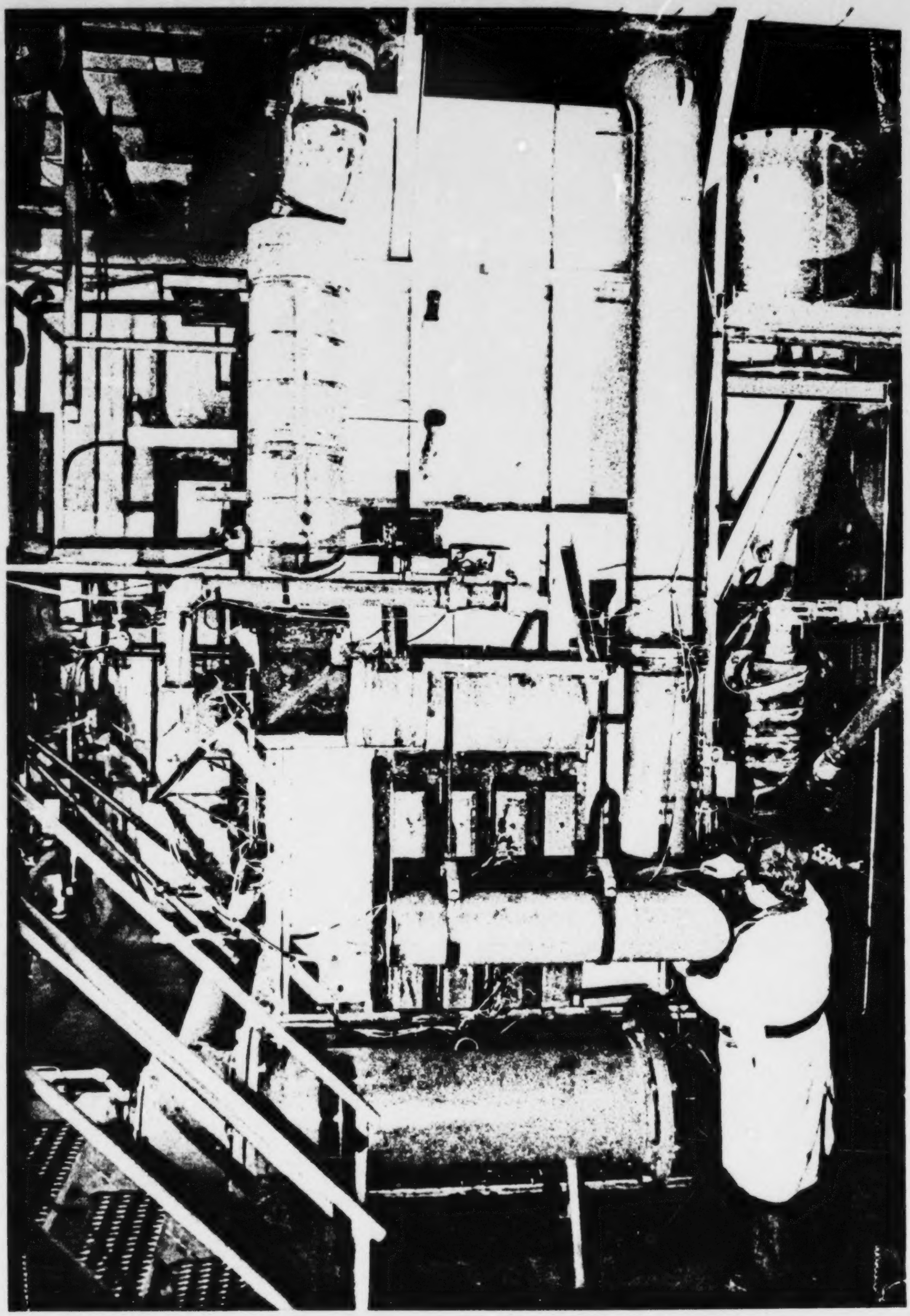


Figure 1. 1 m x 1.6 m fluidized-bed combustor

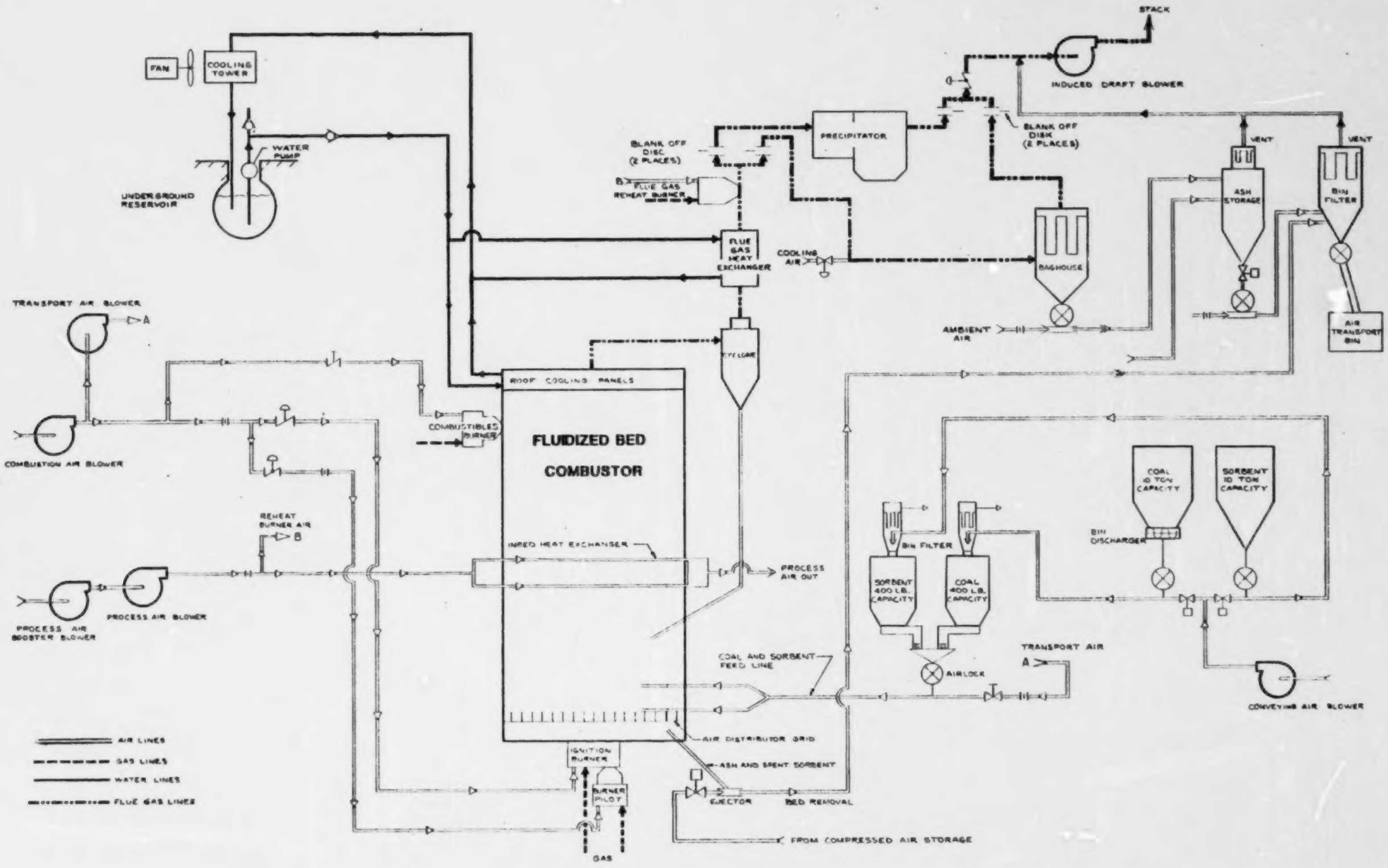

STACK
INDUCED DRAFT BLOWER
FAN
COOLING TOWER
WATER PUMP
UNDERGROUND RESERVOIR
BLANK OFF DISC (2 PLACES)
PRECIPITATOR
BLANK OFF DISK (2 PLACES)
VENT
VENT
B
FLUE GAS REHEAT BURNER
ASH STORAGE
BIN FILTER
FLUE GAS HEAT EXCHANGER
COOLING AIR
BAGHOUSE
TRANSPORT AIR BLOWER
A
AMBIENT AIR
AIR TRANSPORT BIN
CYCLONE
ROOF COOLING PANELS
COMBUSTION AIR BLOWER
COMBUSTIBLES BURNER
FLUIDIZED BED COMBUSTOR
COAL 10 TON CAPACITY
SORBENT 10 TON CAPACITY
BIN DISCHARGER
REHEAT BURNER AIR
B
INBED HEAT EXCHANGER
PROCESS AIR OUT
BIN FILTER
SORBENT 400 LB. CAPACITY
COAL 400 LB. CAPACITY
PROCESS AIR BOOSTER BLOWER
PROCESS AIR BLOWER
TRANSPORT AIR
A
COAL AND SORBENT FEED LINE
AIRLOCK
CONVEYING AIR BLOWER
AIR LINES
GAS LINES
WATER LINES
FLUE GAS LINES
AIR DISTRIBUTOR GRID
IGNITION BURNER
ASH AND SPENT SORBENT
BURNER PILOT
EJECTOR
BED REMOVAL
FROM COMPRESSED AIR STORAGE
GAS

Figure 2. Process schematic for 1 m X 1.6 m fluidized bed test facility

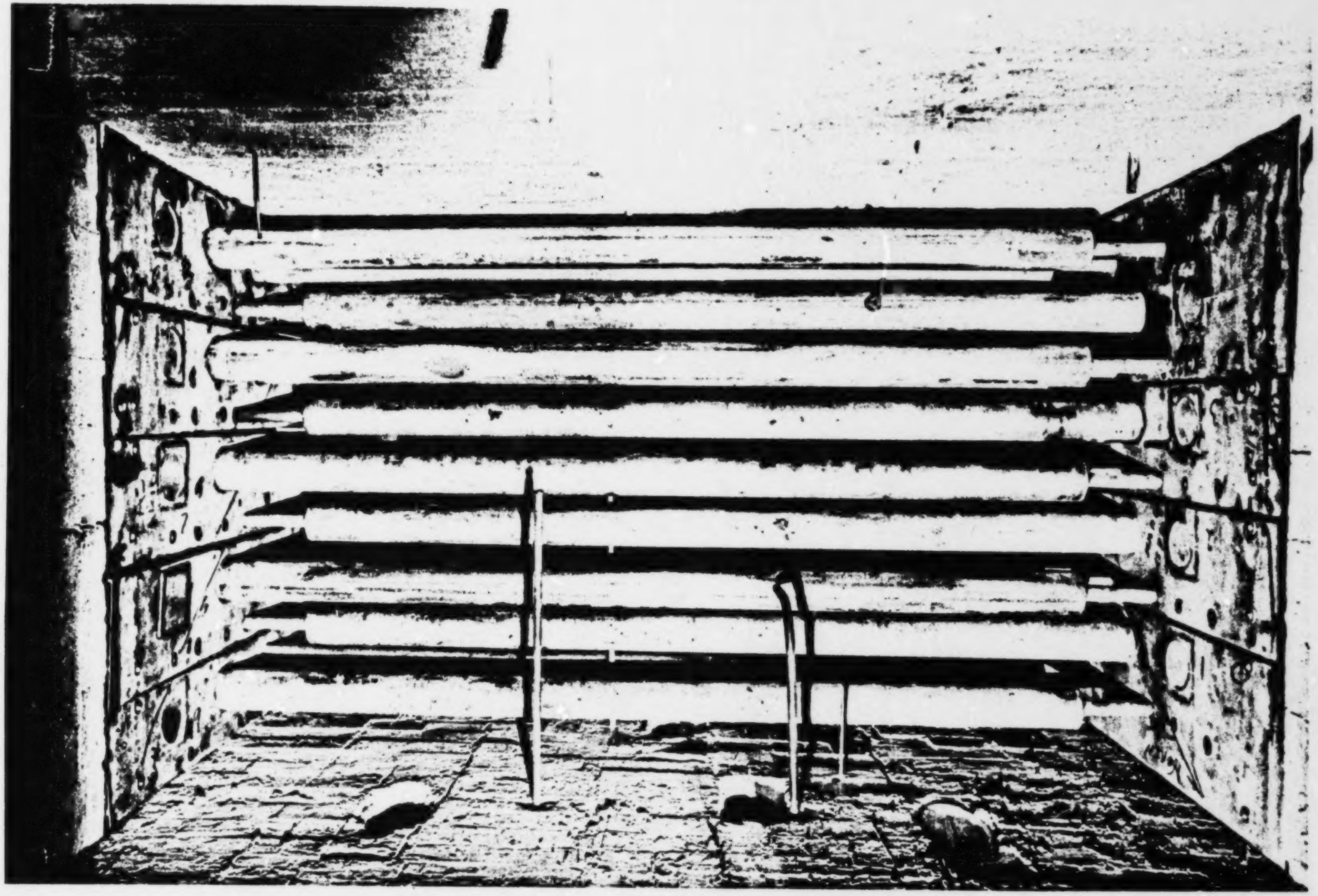
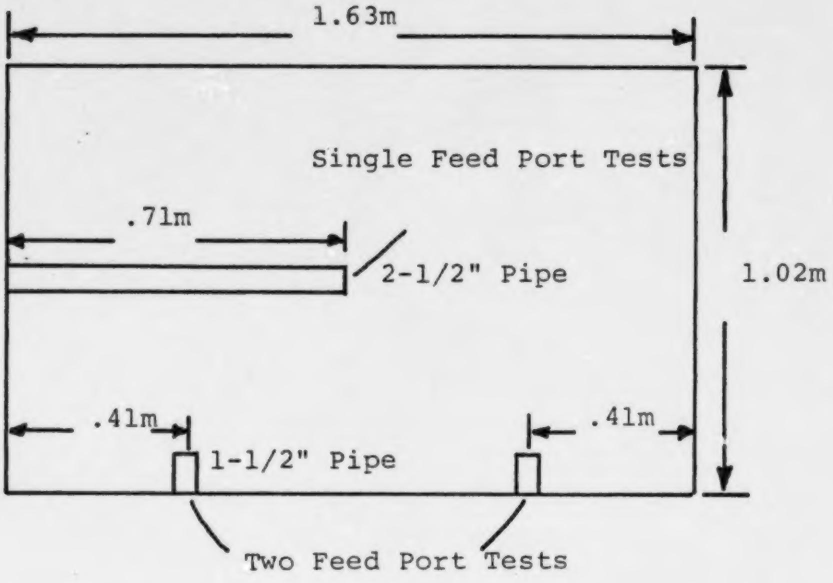


Figure 3. Top view of in-bed tube bundle


1.63m
Single Feed Port Tests
.71m
2-1/2" Pipe
1.02m
.41m
1-1/2" Pipe
.41m
Two Feed Port Tests

Elevation of Ports is 20 cm
above the Distributor Grid

Figure 4. Schematic of feedport locations top view of bed

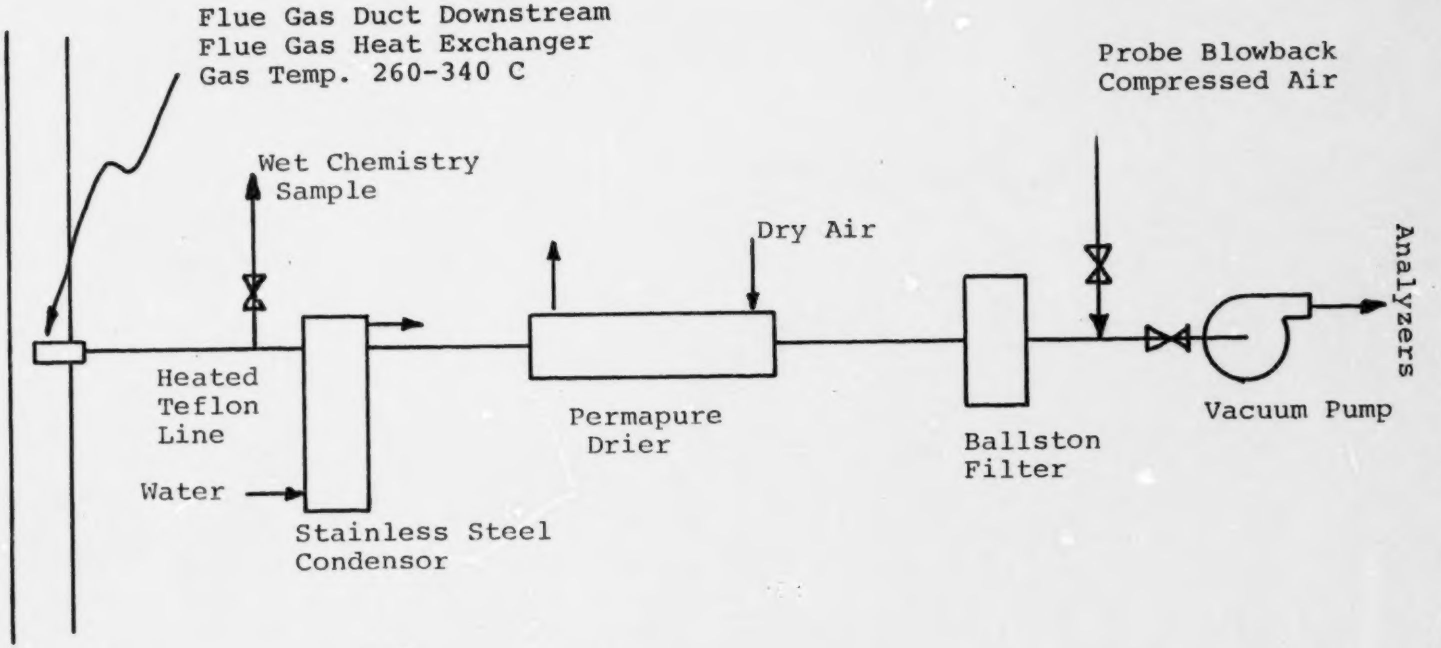

Flue Gas Duct Downstream
Flue Gas Heat Exchanger
Gas Temp. 260-340 C
Wet Chemistry
Sample
Heated
Teflon
Line
Water
Stainless Steel
Condensor
Permapure
Drier
Dry Air
Ballston
Filter
Probe Blowback
Compressed Air
Vacuum Pump
Analyzers

Figure 5. Flue gas sampling system

Mass mean sorbent particle size (microns)	800			1300		
Gas velocity (m/sec)	0.9		1.4	1.4	1.8	
Bed depth (m)	0.9	1.4	0.9	1.4	0.9	1.4
Residence time (sec)	1.0	1.5	0.7	1.0	0.5	0.75
Ca/S	2.0*, 2.75*	2.0, 2.75	2.0, 2.75	2.0, 2.75	2.0, 2.75*	2.0, 2.75*

* The 4 tests marked with an asterisk were replicated, to yield a total of 16 tests.

Conditions for all 16 tests:
- Illinois #6 coal (3.5 % S)
- Greer limestone
- Bed temperature 844 °C
- Excess oxygen 5% (30% excess air)
- Two coal/sorbent feed ports
- No recycle

Figure 6. Test matrix for Tests 13-28 (effects of sorbent size and gas residence time)

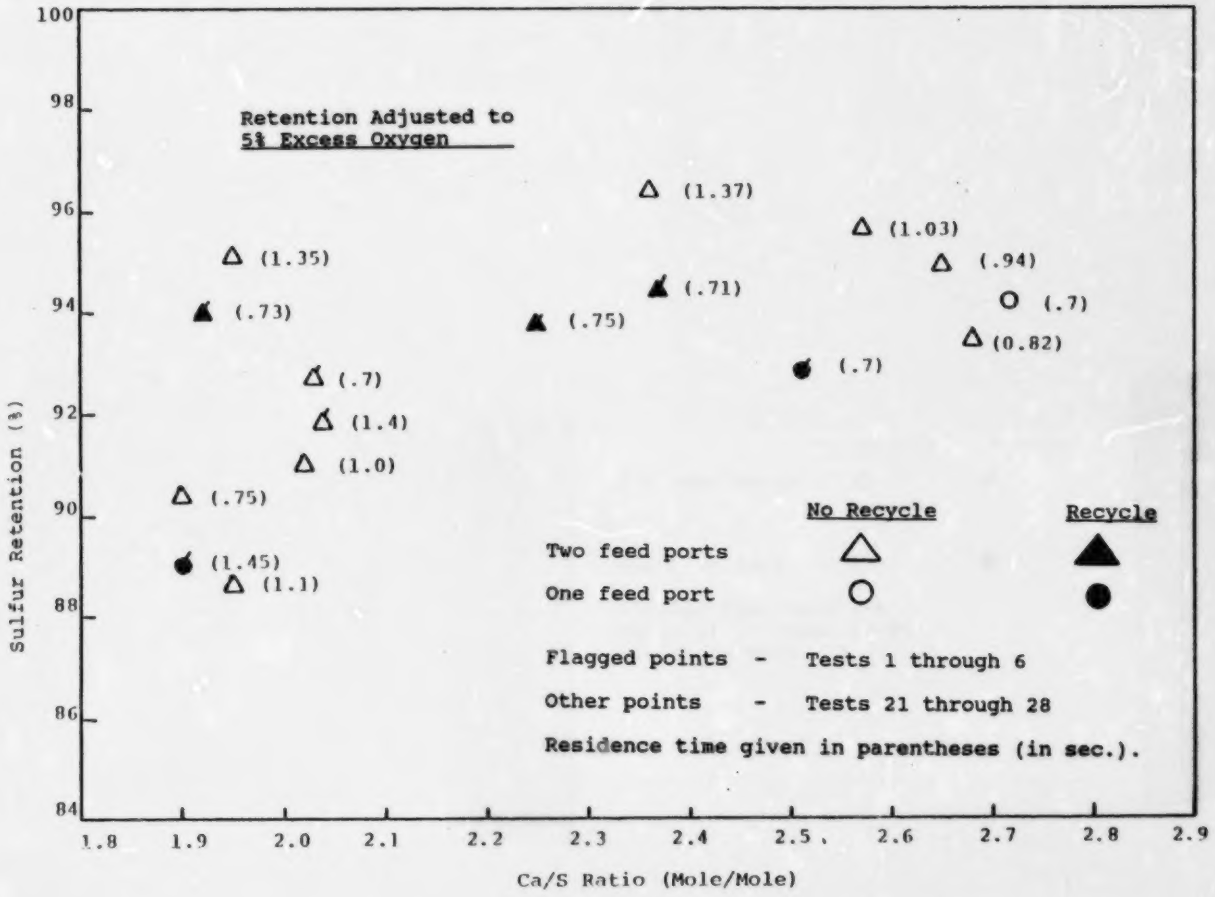


Figure 7. Sulfur retention vs. Ca/S molar ratio
(-8 mesh limestone feed, 800 micron mass mean)

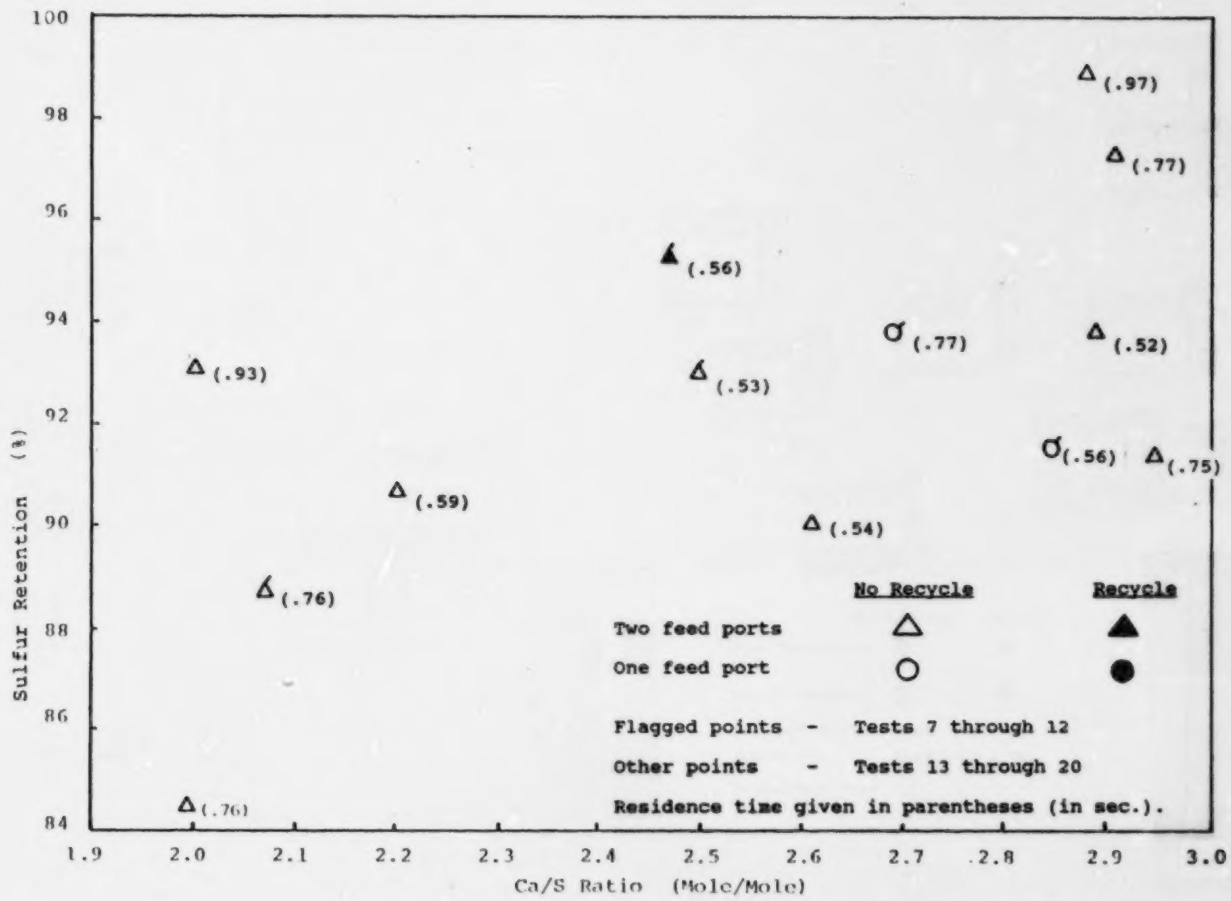


Figure 8. Sulfur retention vs. Ca/S molar ratio
(4 x 30 mesh limestone feed, 1300 micron mass mean)

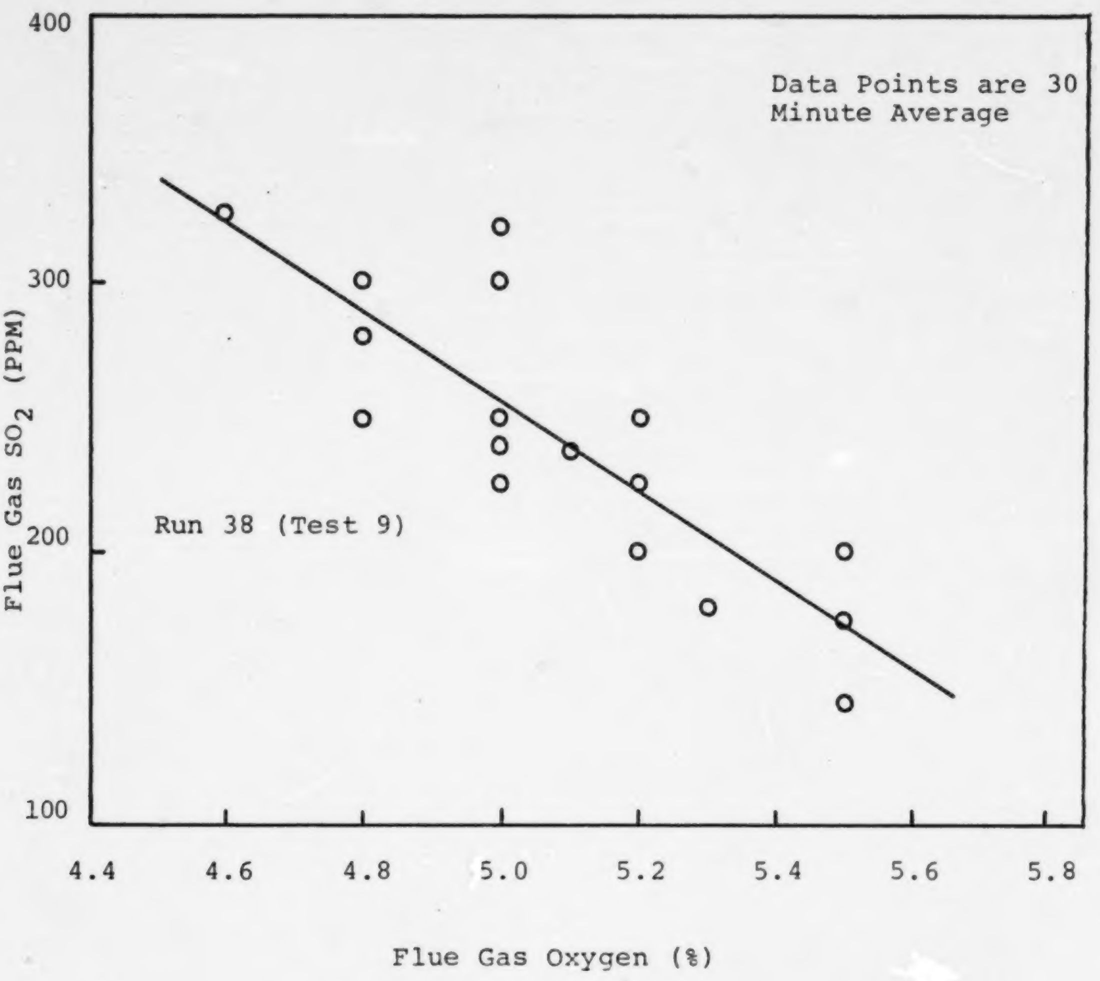


Figure 9. SO_2 emissions vs. flue gas oxygen

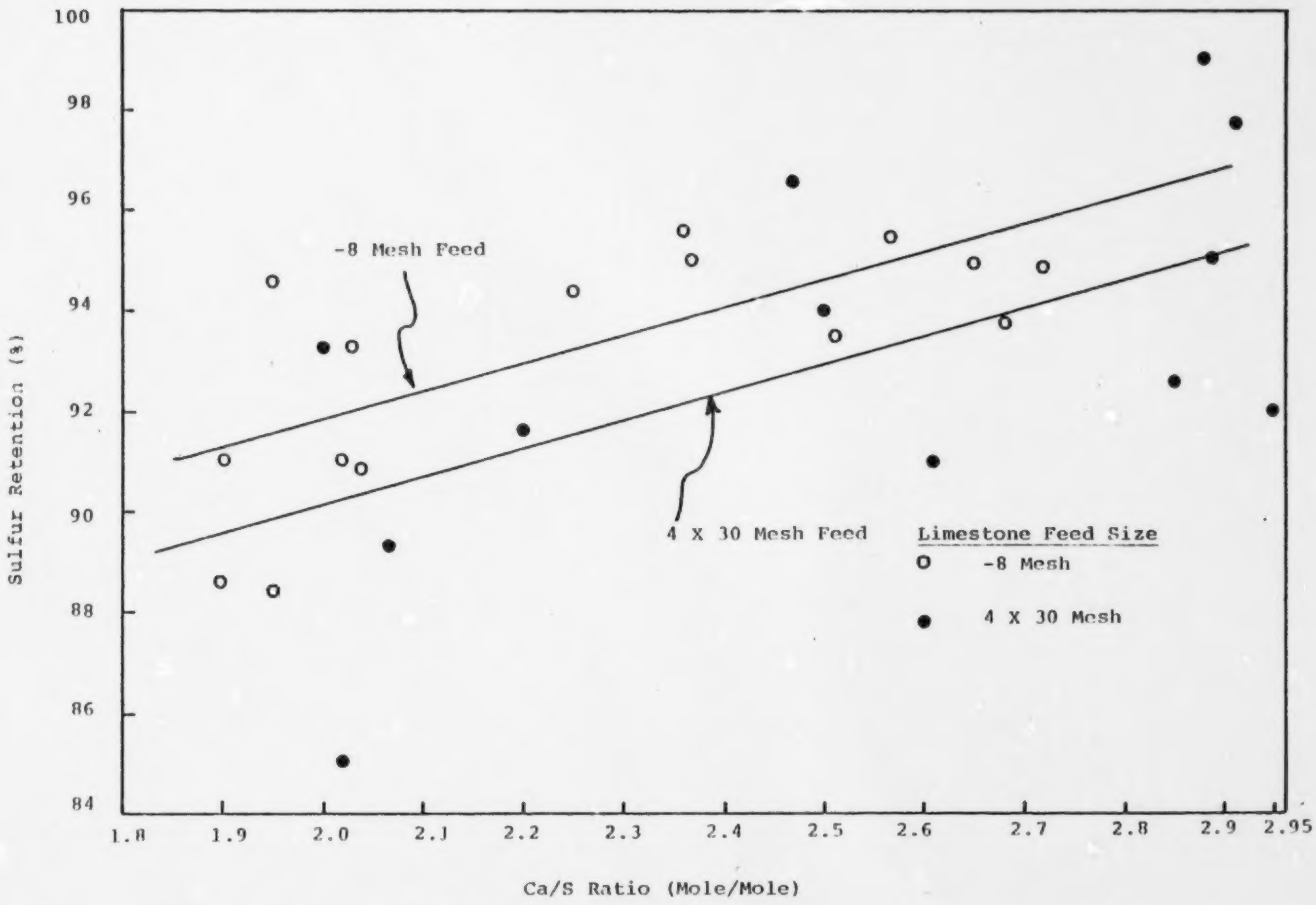


Figure 10. Effect of sorbent feed rate on sulfur retention (adjusted to a gas residence time of 1.0 sec)

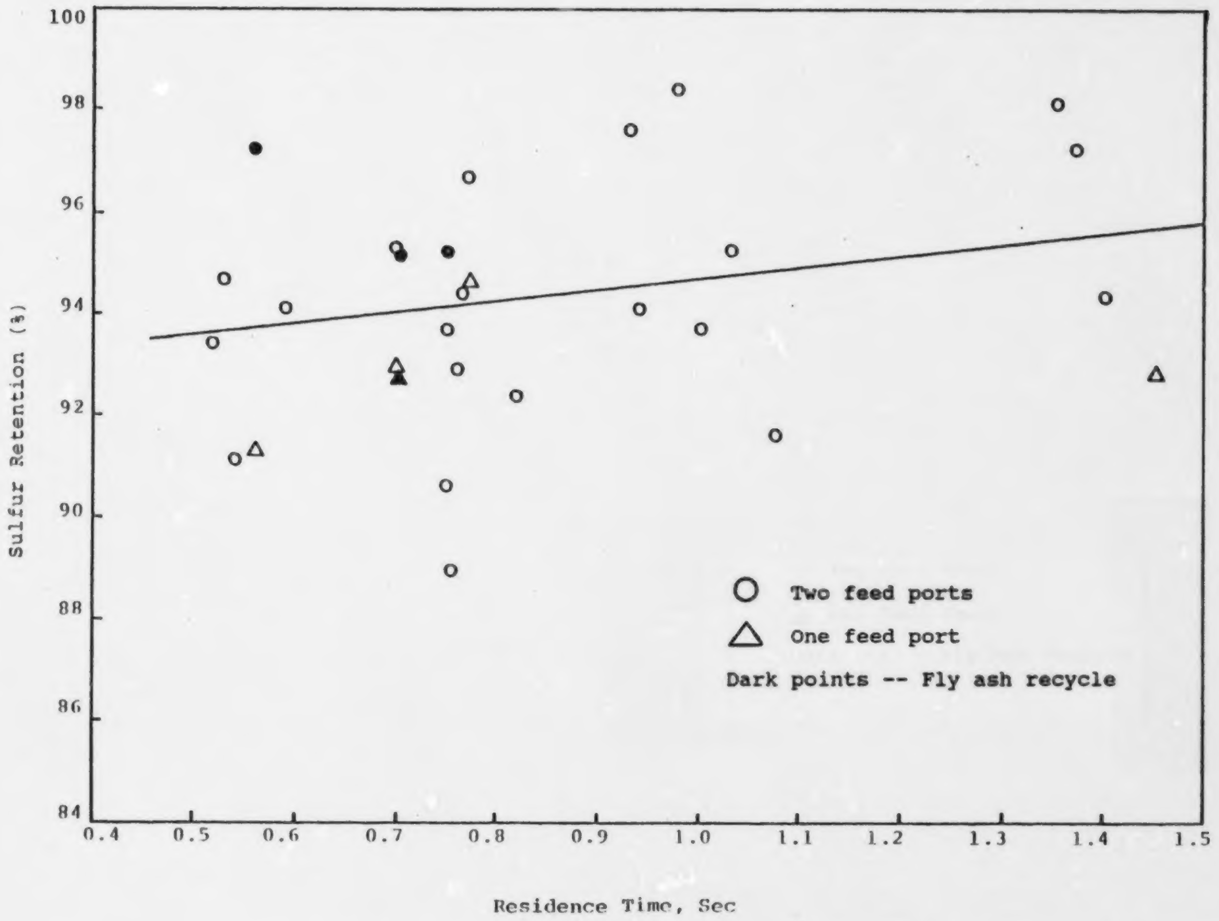


Figure 11. Effect of gas residence time on sulfur retention (adjusted to Ca/S ratio of 2.5, limestone feed size of -8 mesh)

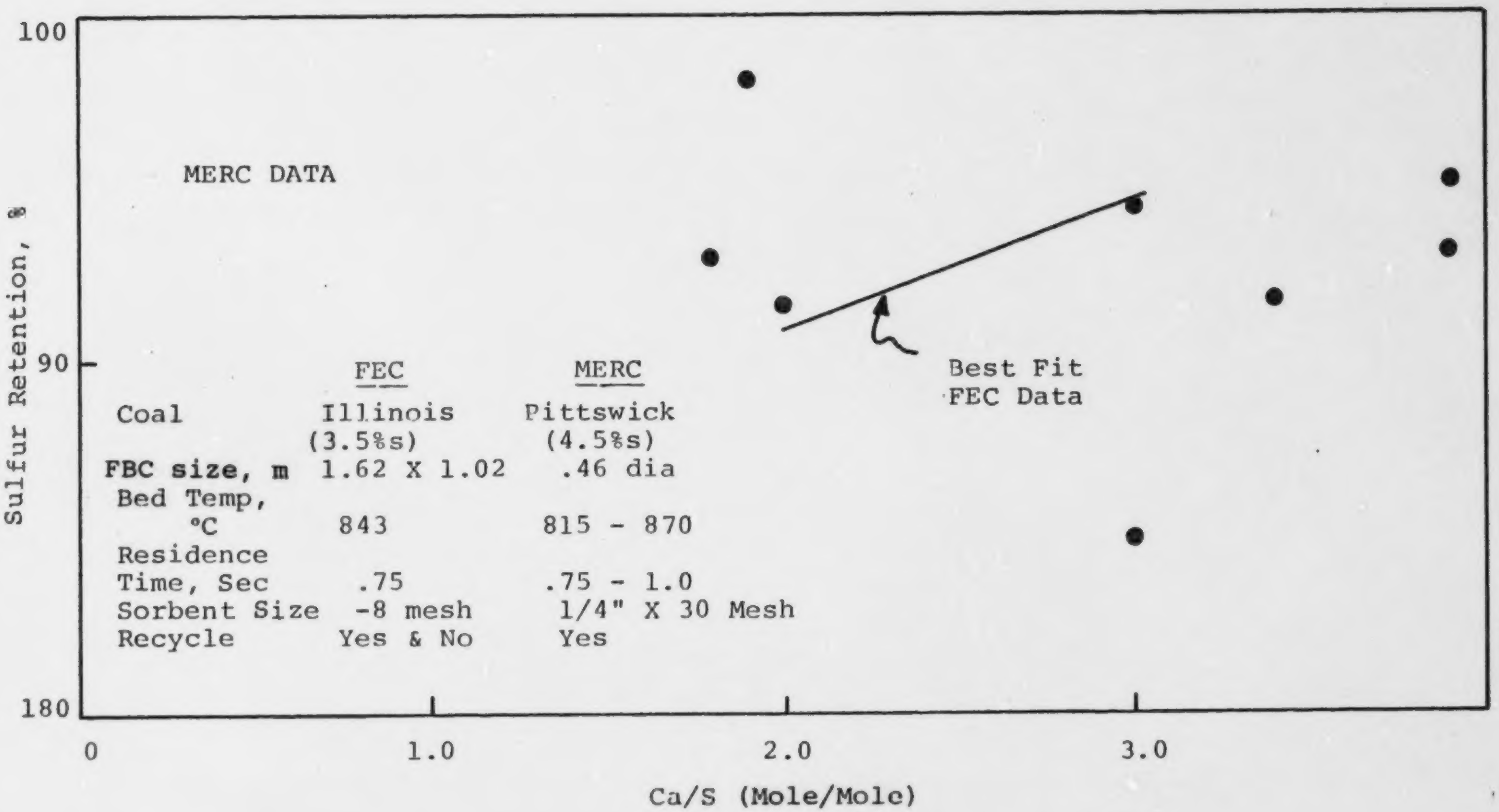


Figure 12. Comparison of sulfur retention data with Greer limestone: data obtained by FluiDyne Engineering Corp. (FEC) in this study vs. data obtained by Morgantown Energy Research Center (MERC)

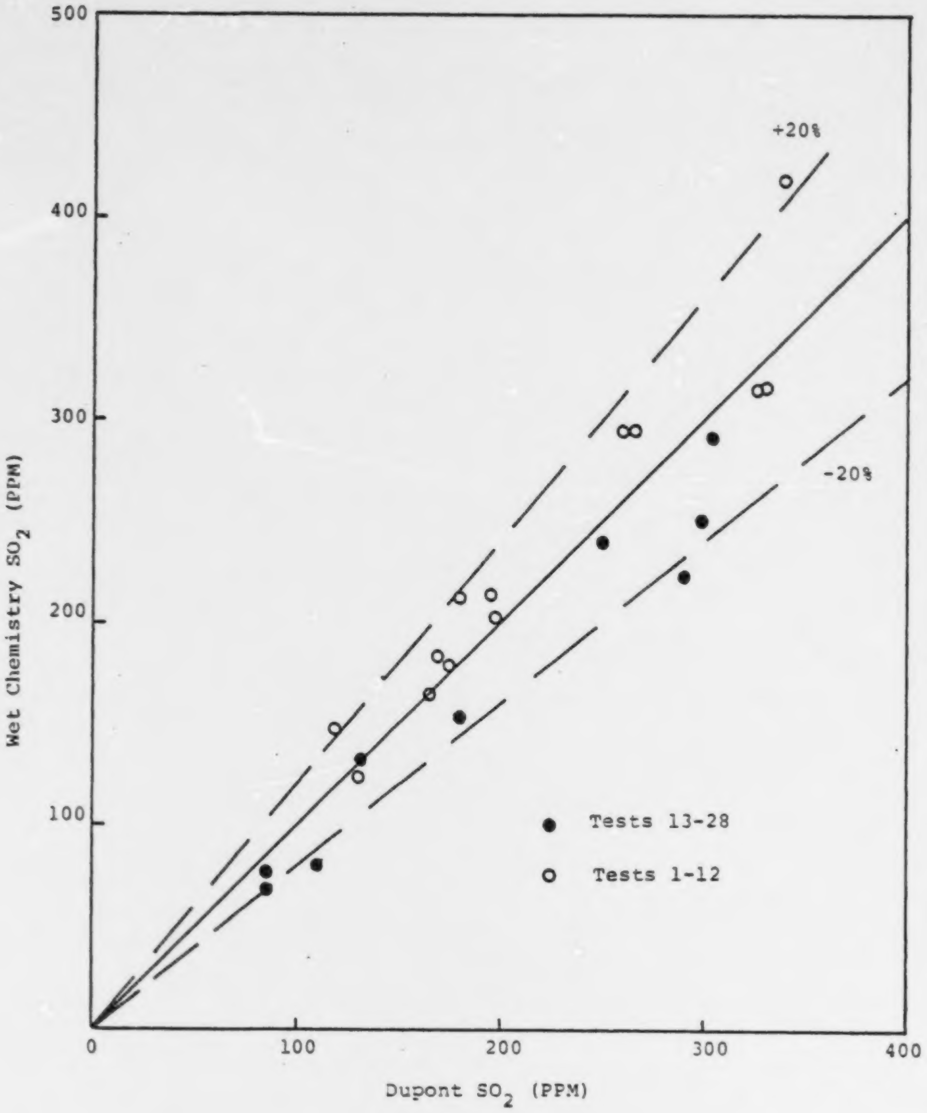


Figure 13. Wet chemistry confirmation of Dupont SO_2 readings

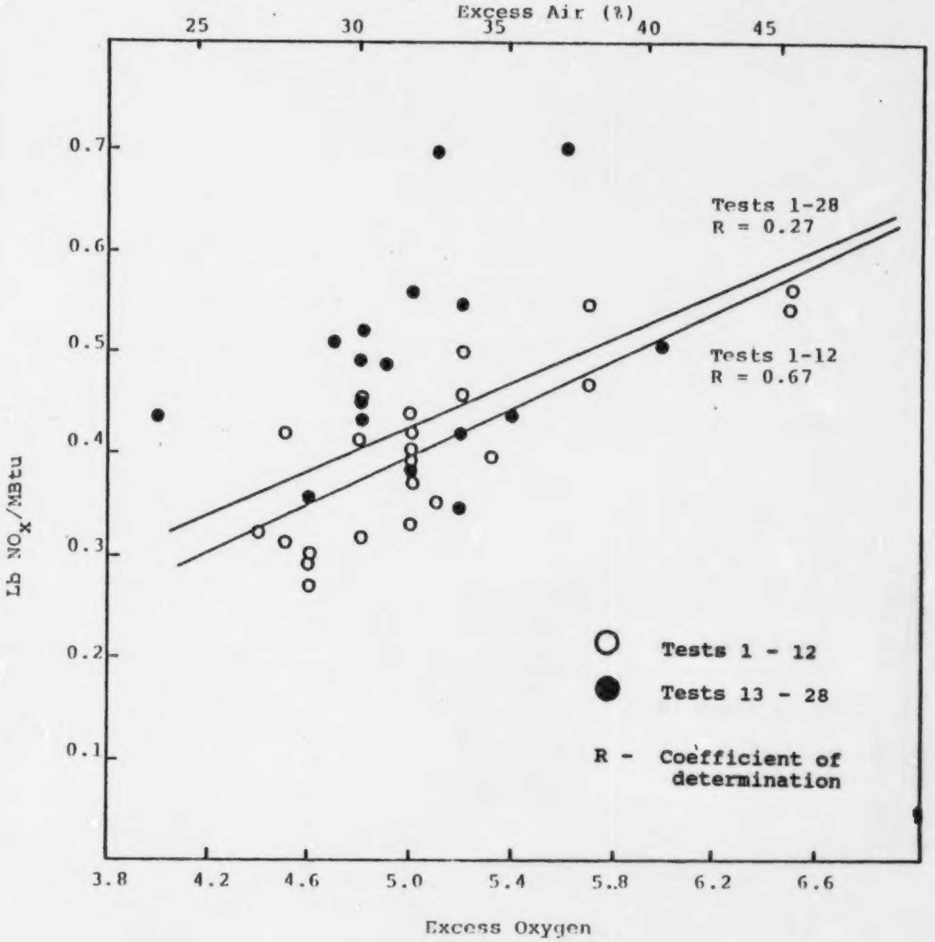


Figure 14. NO$_x$ emissions vs. excess oxygen

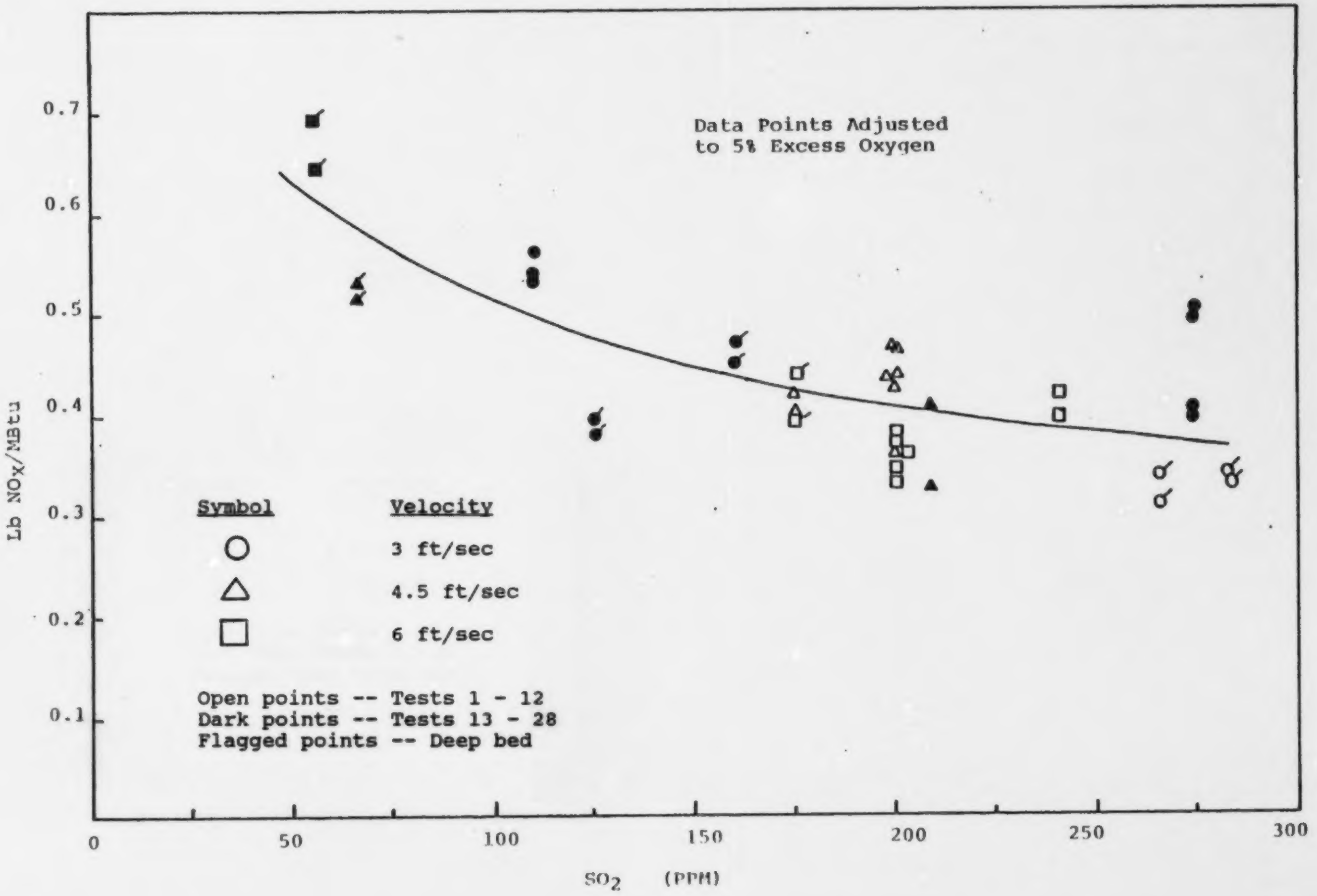


Figure 15. NO_x emissions (adjusted to 5% excess oxygen) vs. SO_2

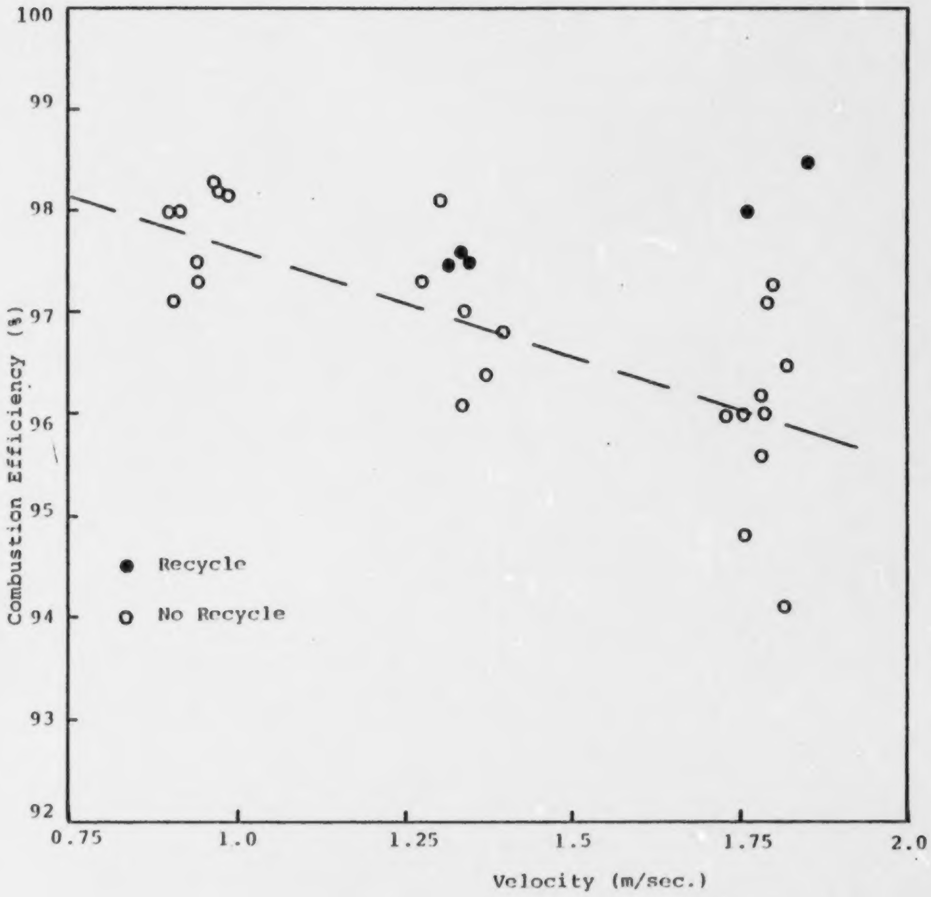


Figure 16. Combustion efficiency vs. velocity

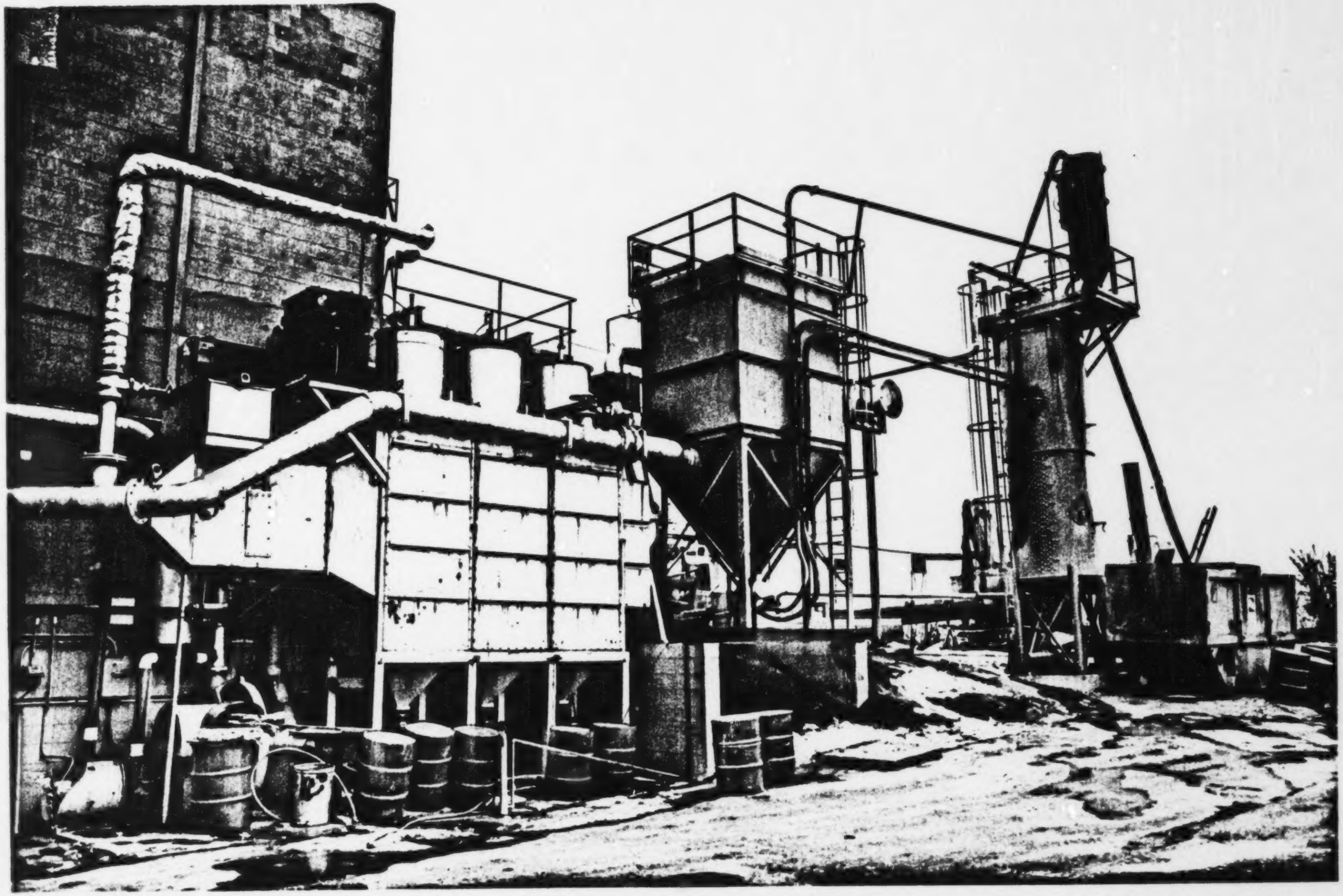


Figure 17. Flue gas cleanup equipment on fluidized-bed combustor. (Left to right: electrostatic precipitator, baghouse, ash storage silo.)

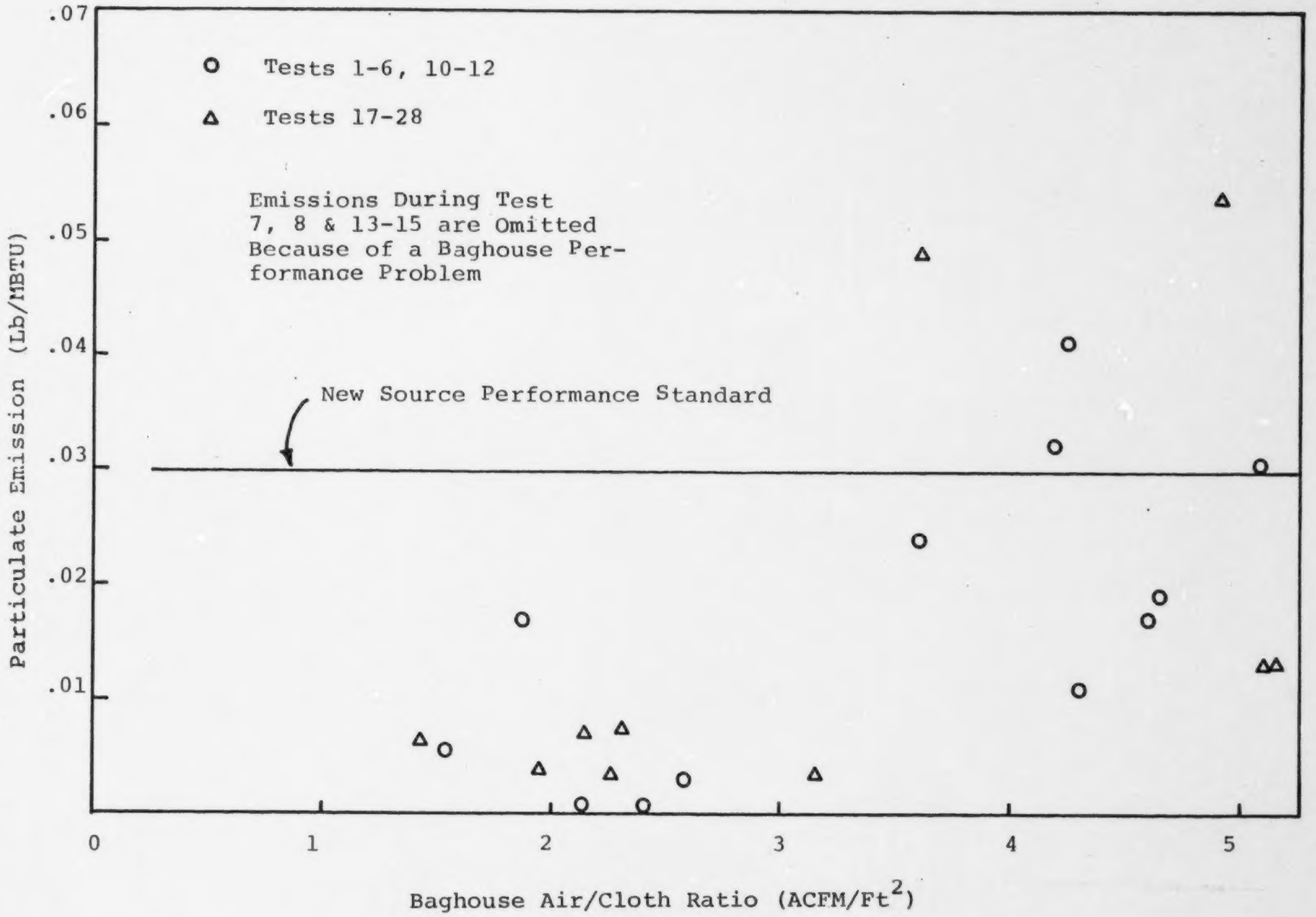


Figure 18. Baghouse particulate emission

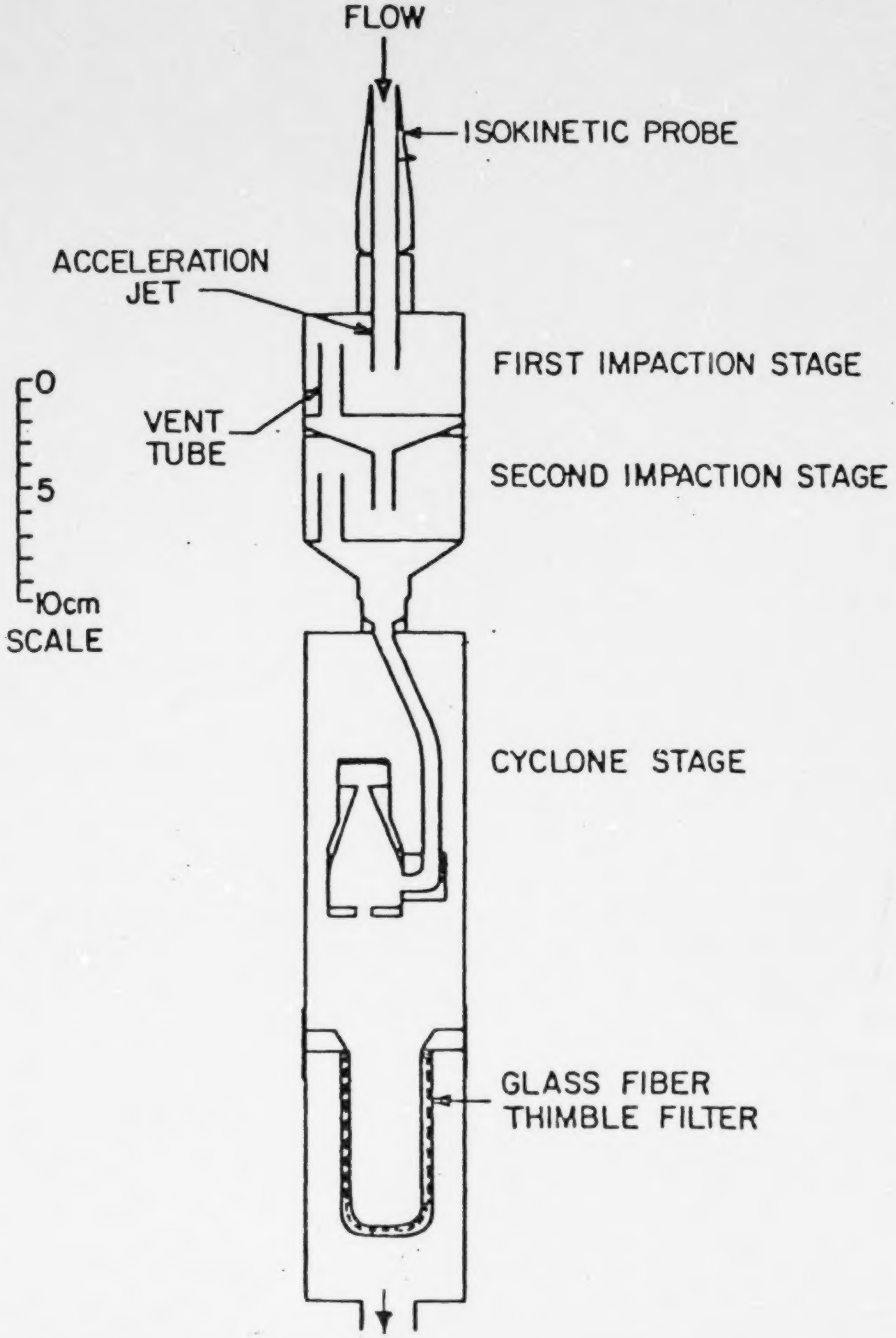

FLOW
ISOKINETIC PROBE
ACCELERATION JET
FIRST IMPACTION STAGE
VENT TUBE
SECOND IMPACTION STAGE
0
5
10cm
SCALE
CYCLONE STAGE
GLASS FIBER THIMBLE FILTER

Figure 19. Schematic of the Andersen model HCSS high grain loading impactor

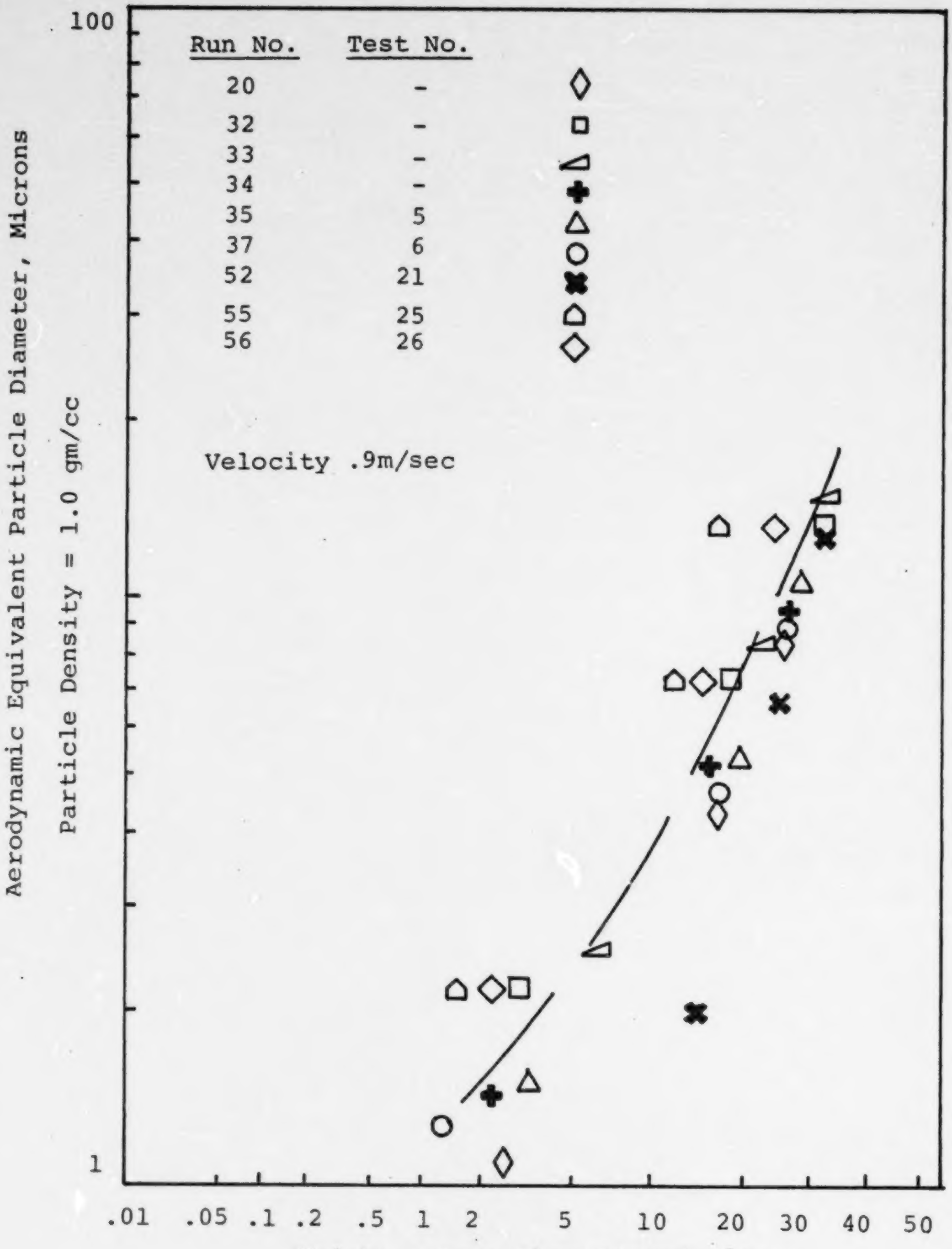


Figure 20. Cascade impactor data acquired upstream of the baghouse (.9 m/sec)

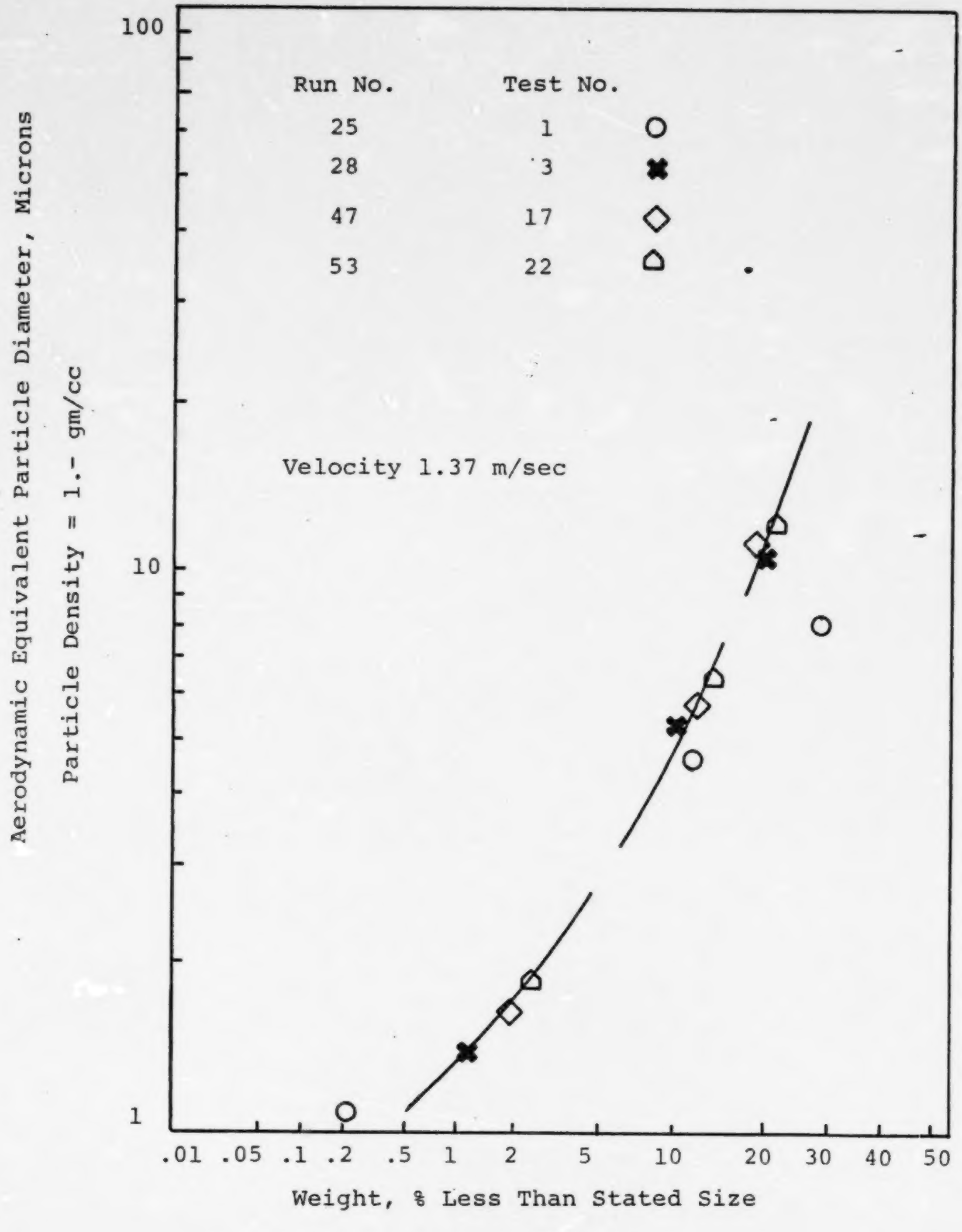


Figure 21. Cascade impactor data acquired upstream of the baghouse, superficial velocity 1.37 m/sec

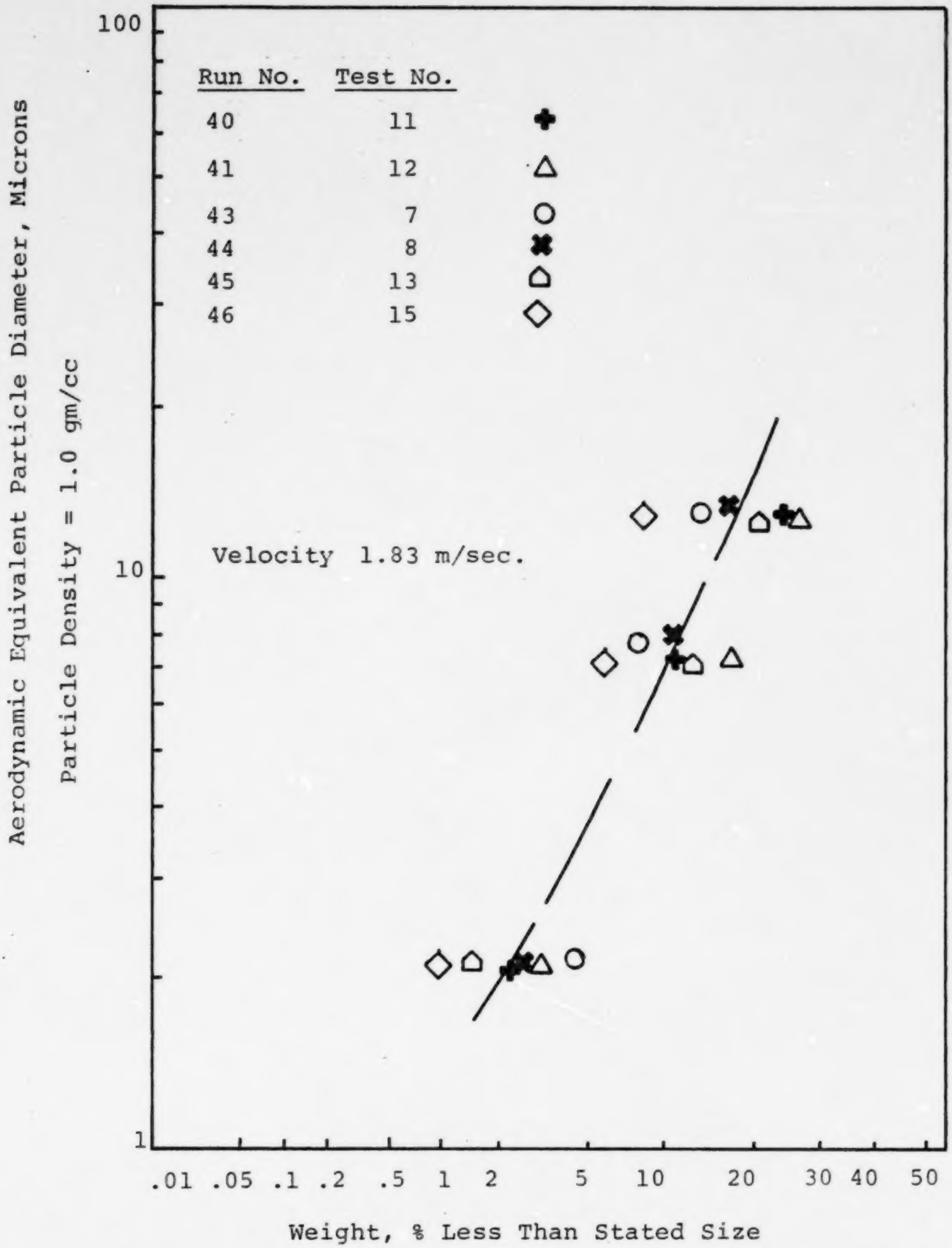


Figure 22. Cascade impactor data acquired upstream of the baghouse - superficial velocity 1.83 m/sec.

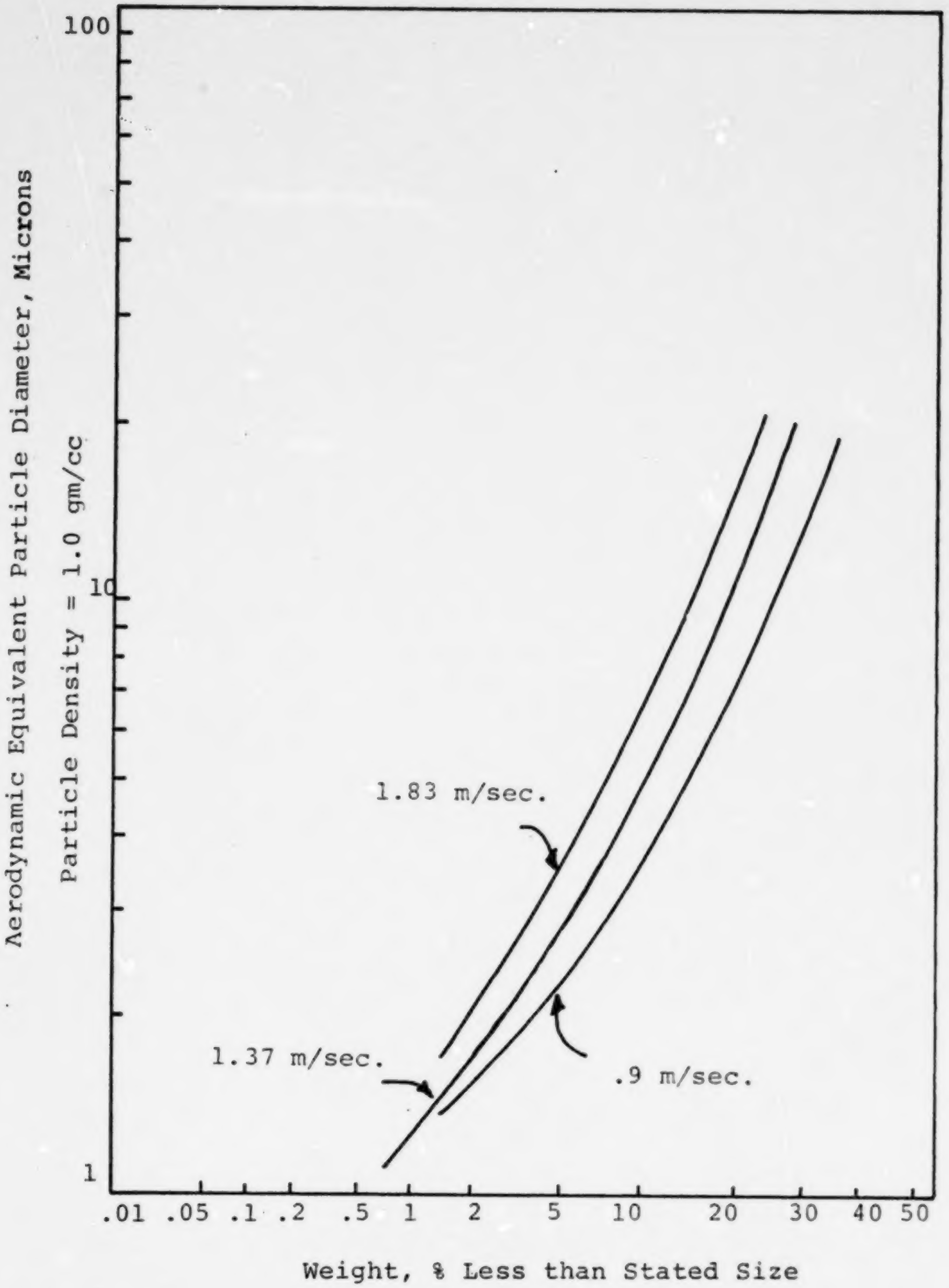


Figure 23. Cascade impactor data acquired upstream of the baghouse - trend with velocity

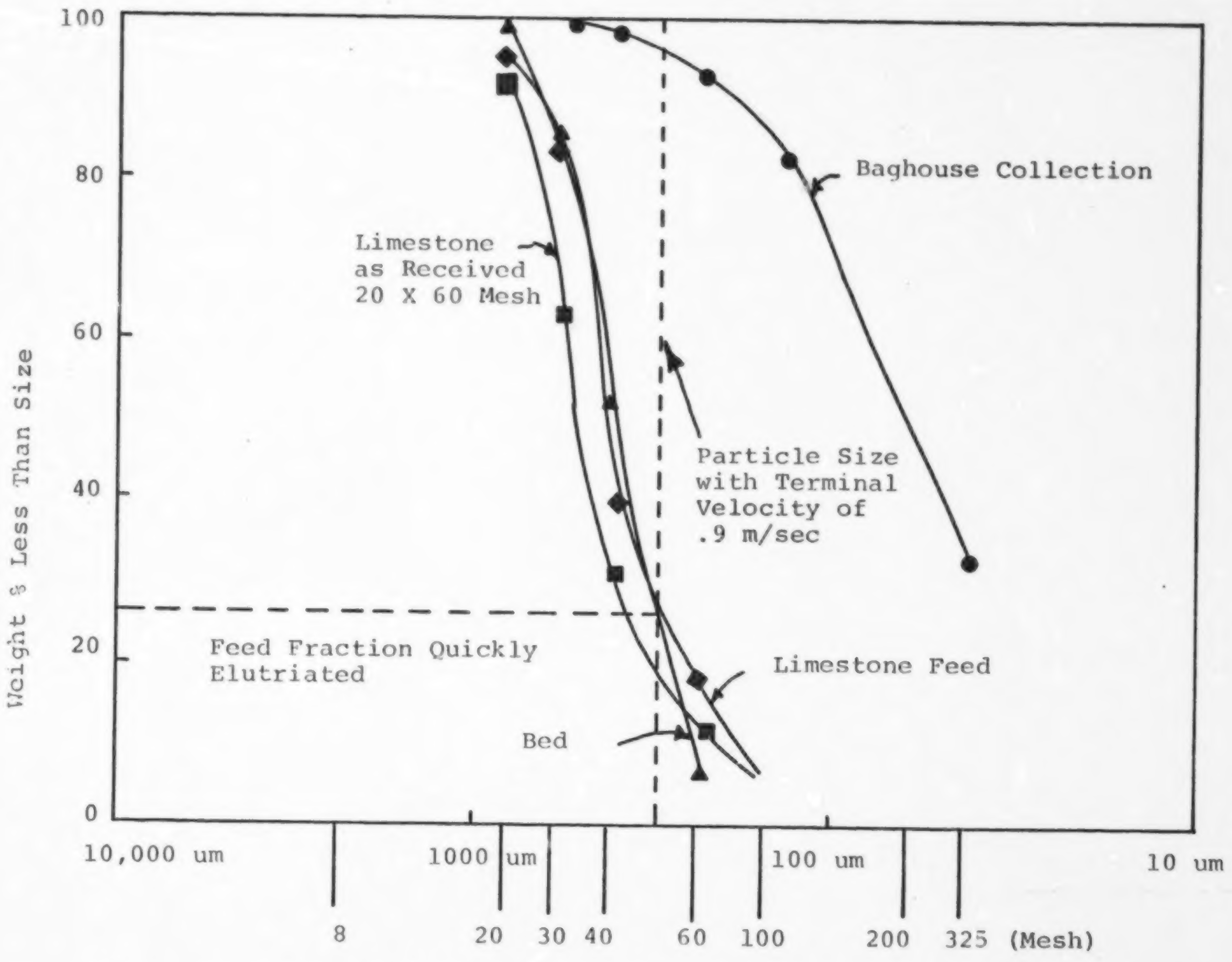


Figure 24. Size distribution of the feed limestone, bed and elutriation in runs with 20 x 60 mesh Calcium Carbonate Co. limestone

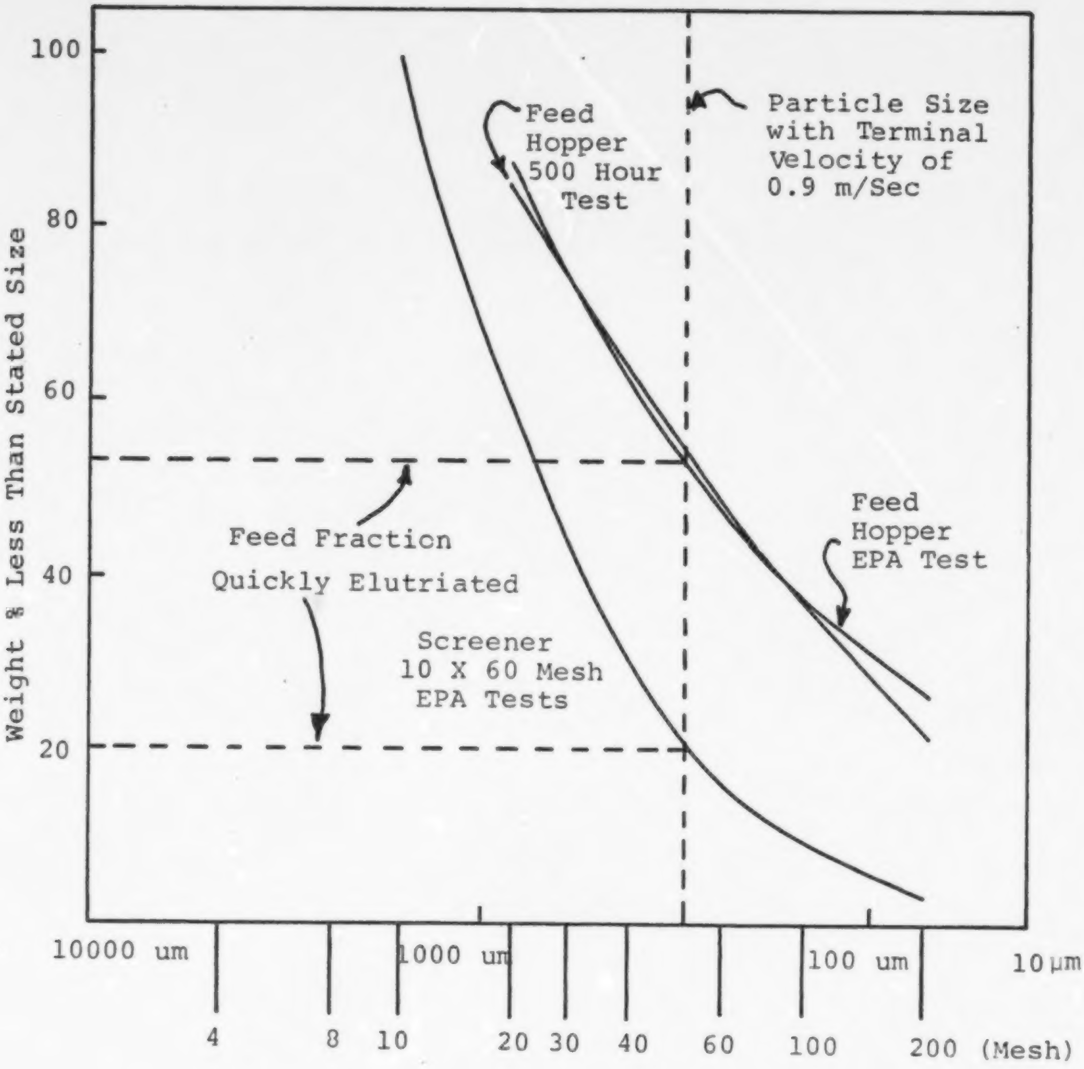


Figure 25. The effect of attrition during pneumatic handling on the feed size distribution of Lundeen dolomite

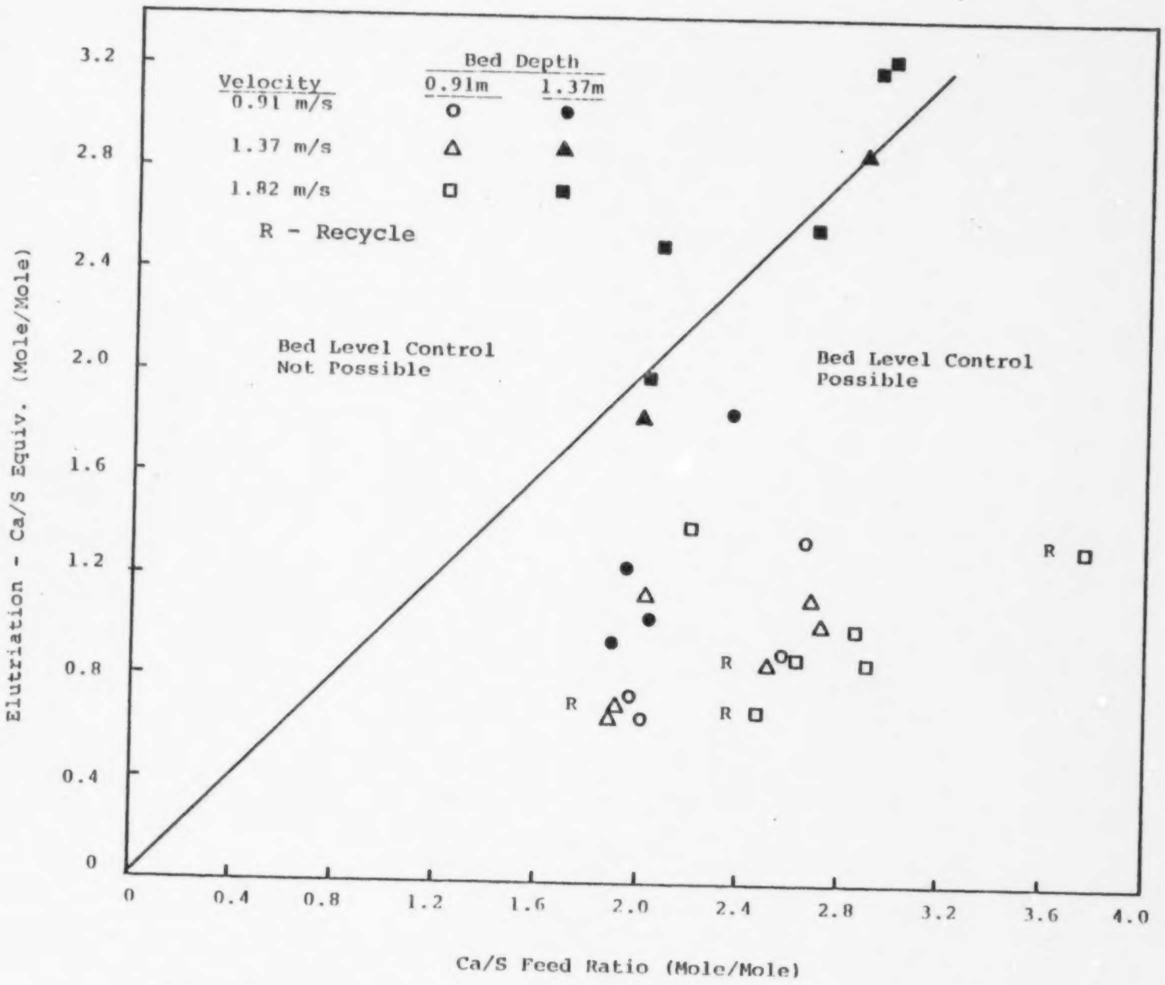


Figure 26. The effect of operating conditions on Greer limestone elutriation

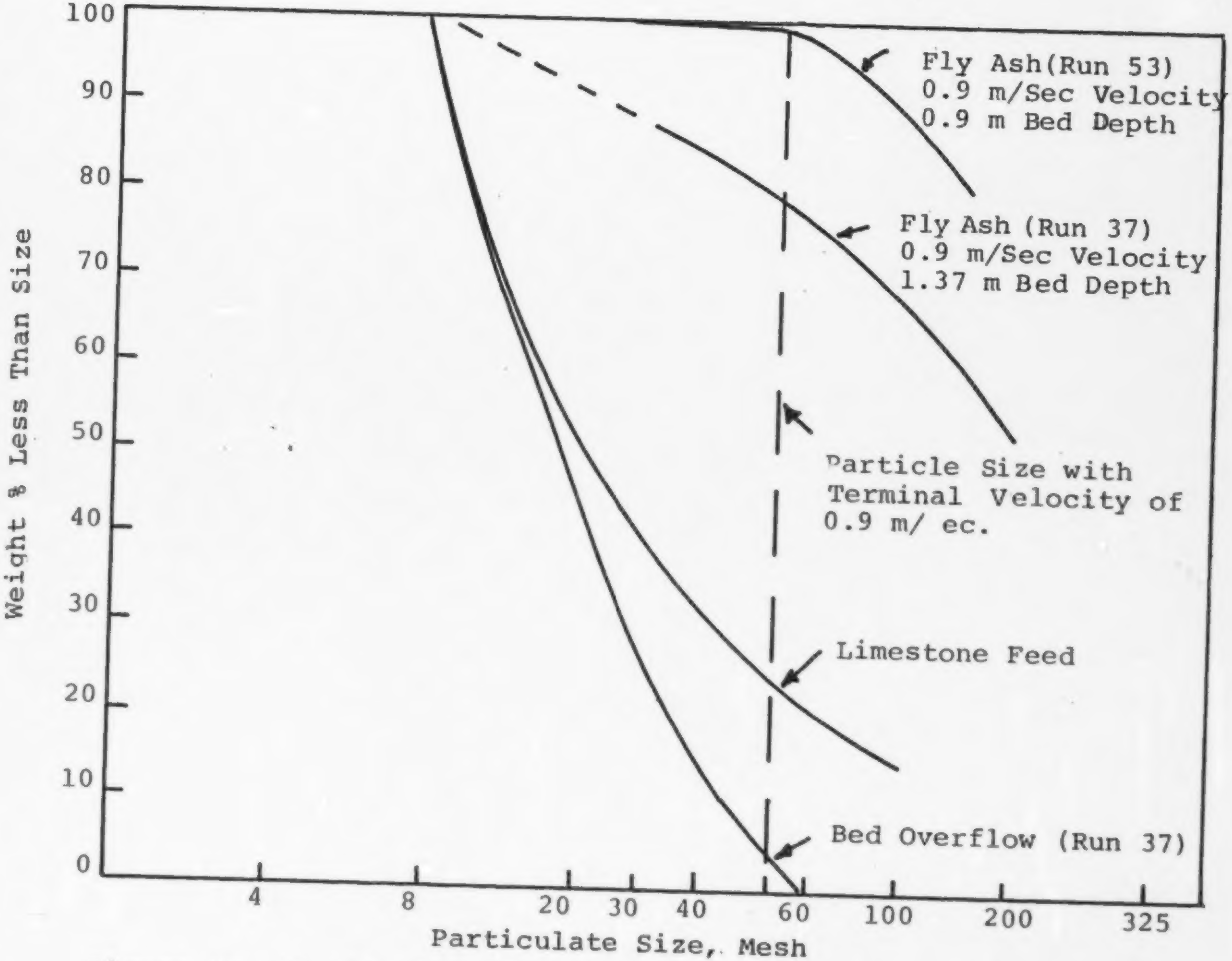


Figure 27. Elutriation size distribution vs. bed height -8 mesh limestone

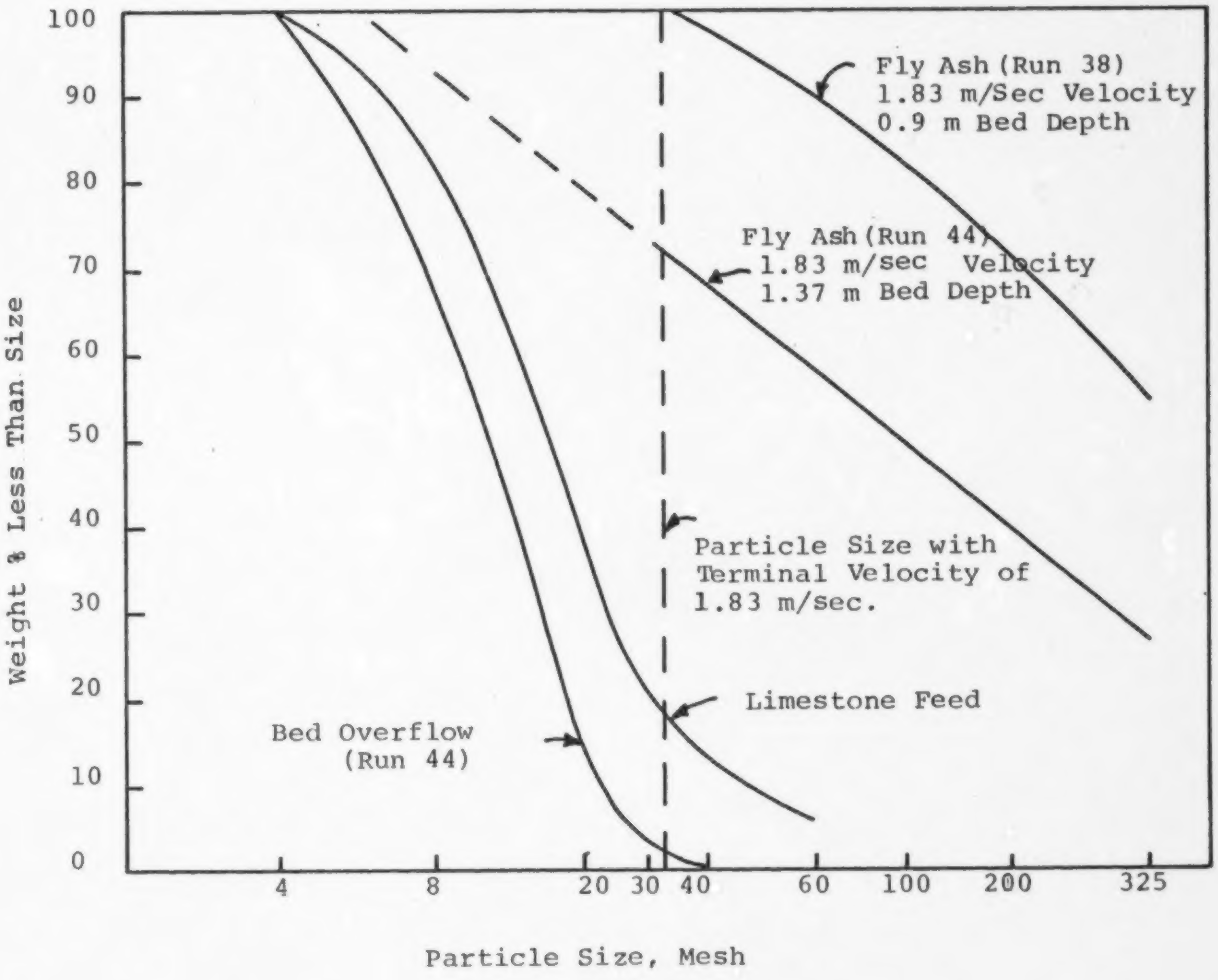


Figure 28. Elutriation size distribution vs. bed height for a 40 x 30 mesh limestone feed

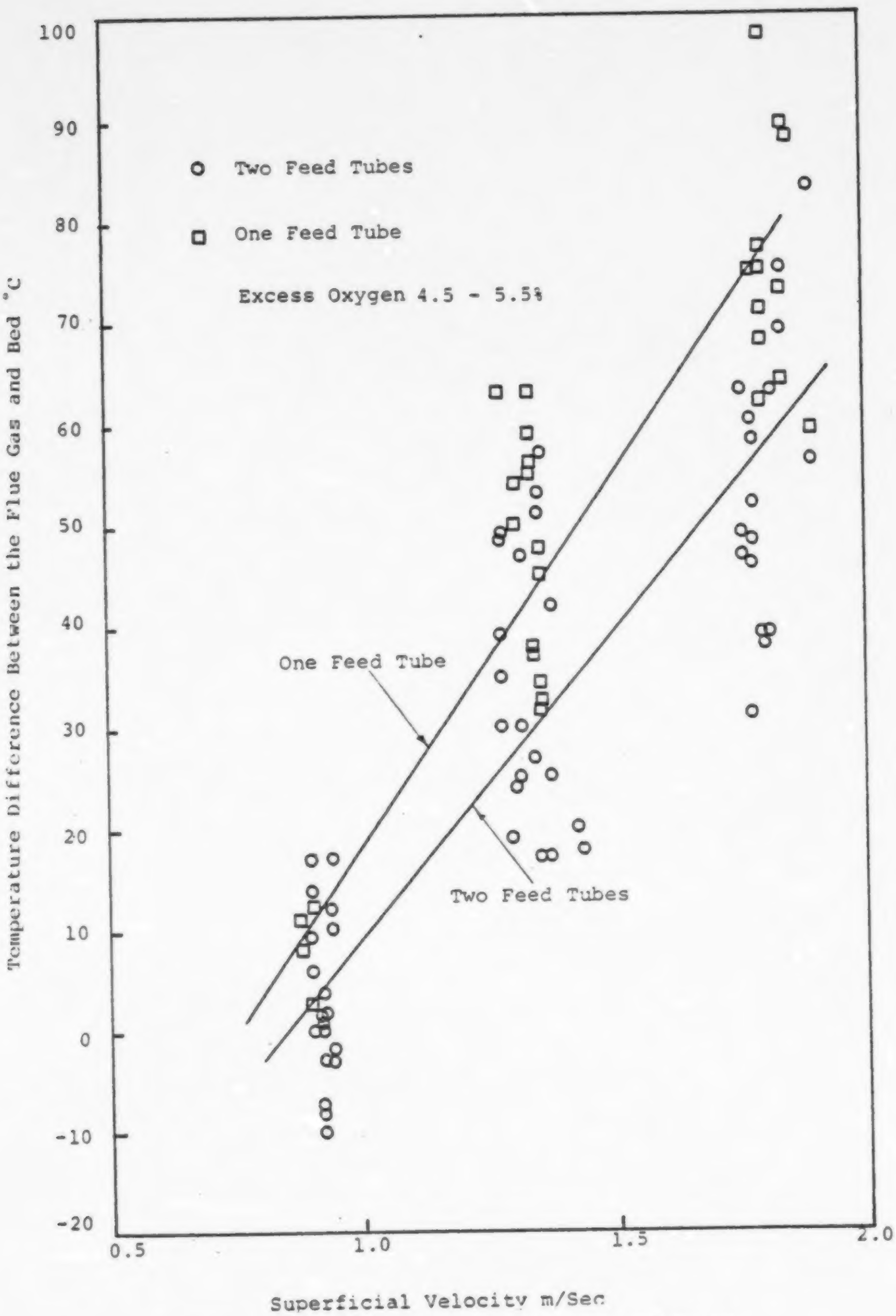


Figure 29. Above bed combustion vs. velocity

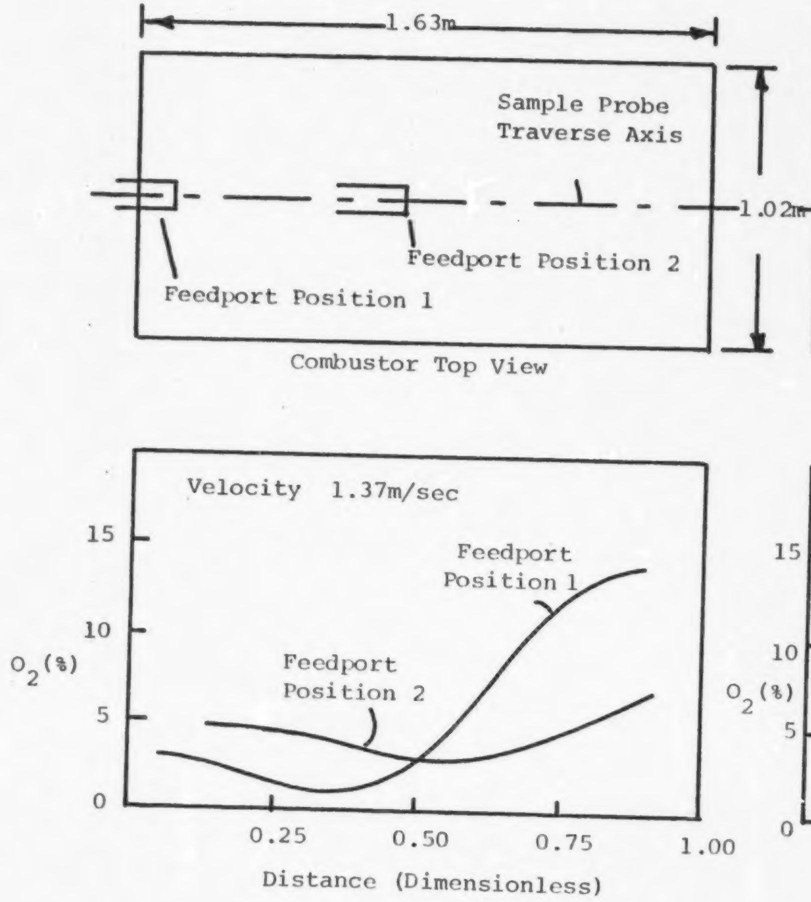


Figure 30. Freeboard surveys - effect of single feedport location

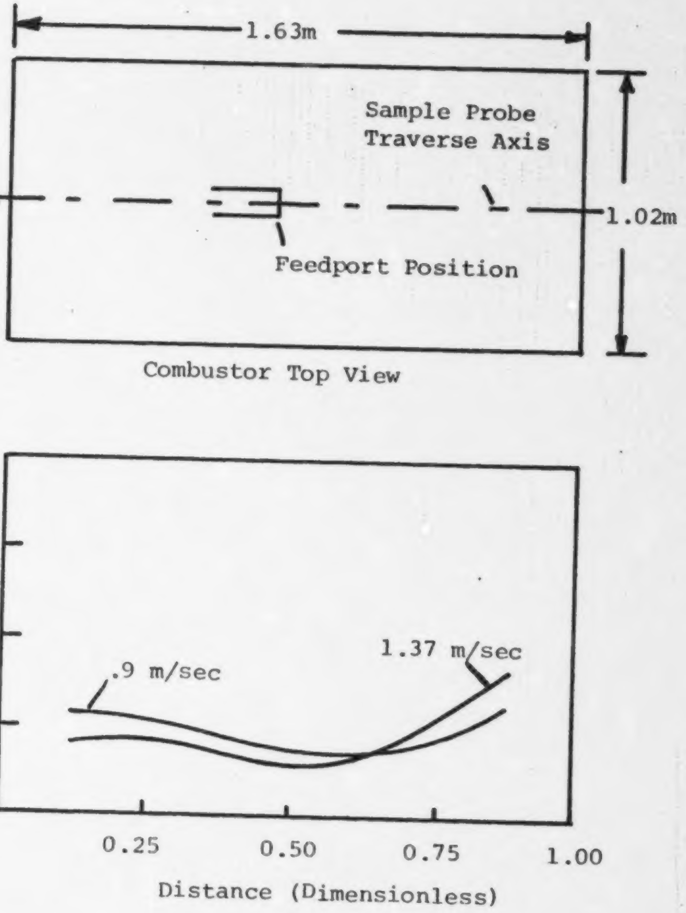


Figure 31. Freeboard surveys - effect of velocity with a single feed tube

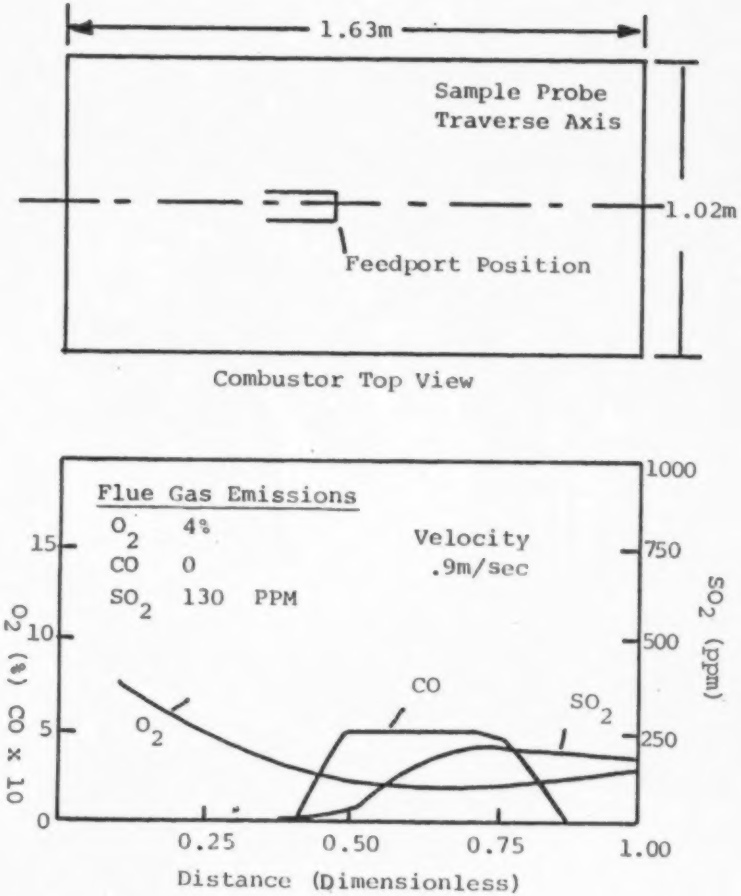


Figure 32. Freeboard surveys - emmision profile single feed tube

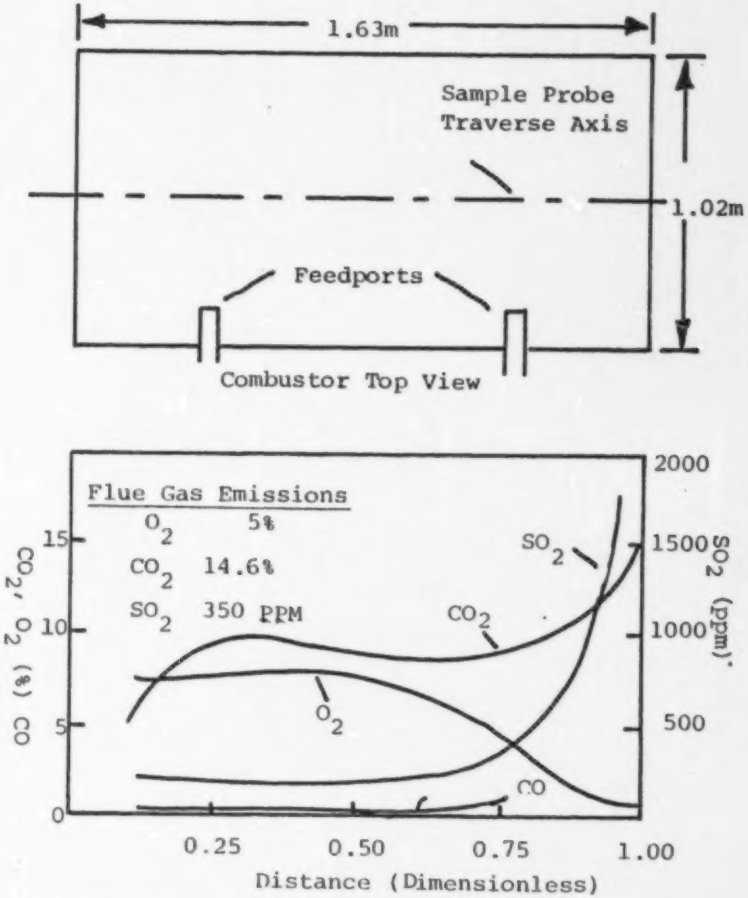
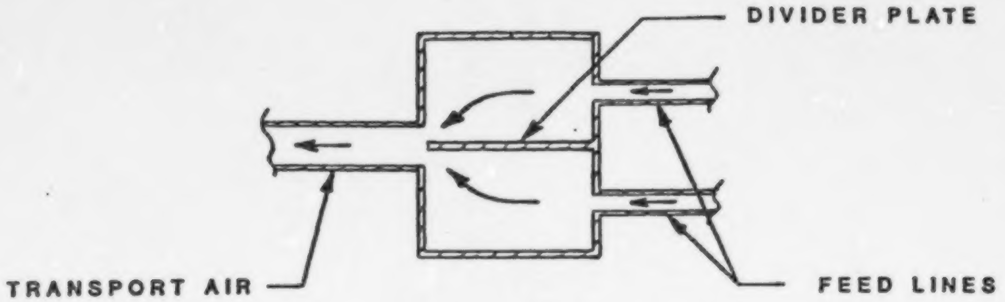


Figure 34. Freeboard surveys - emission profile plate feed splitter


DIVIDER PLATE
TRANSPORT AIR
FEED LINES

SECTION A-A

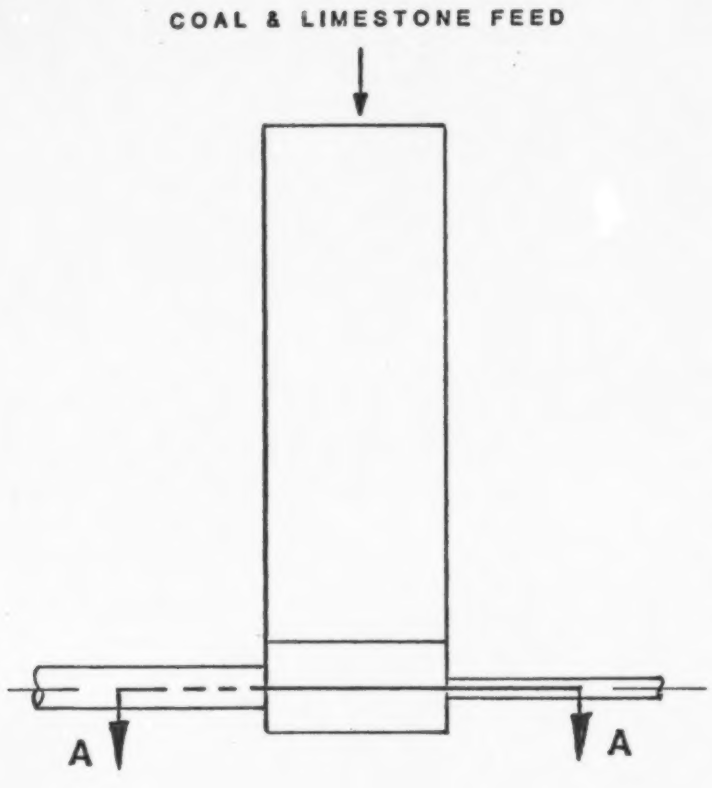

COAL & LIMESTONE FEED
A
A

Figure 33. Feed splitter design using a divider plate

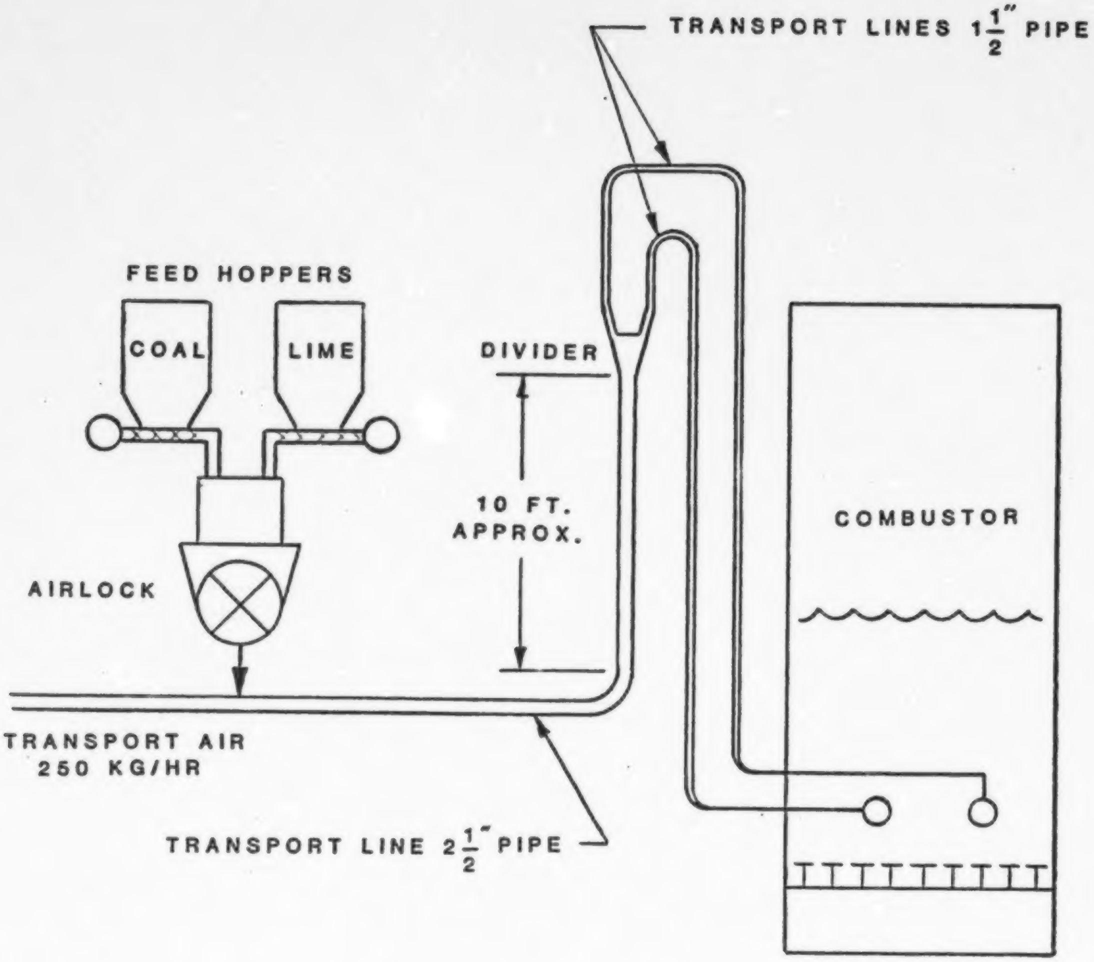

TRANSPORT LINES 1$\frac{1}{2}$" PIPE
FEED HOPPERS
COAL
LIME
DIVIDER
10 FT. APPROX.
COMBUSTOR
AIRLOCK
TRANSPORT AIR 250 KG/HR
TRANSPORT LINE 2$\frac{1}{2}$" PIPE

Figure 35. Schematic of the "upflow" feed divider

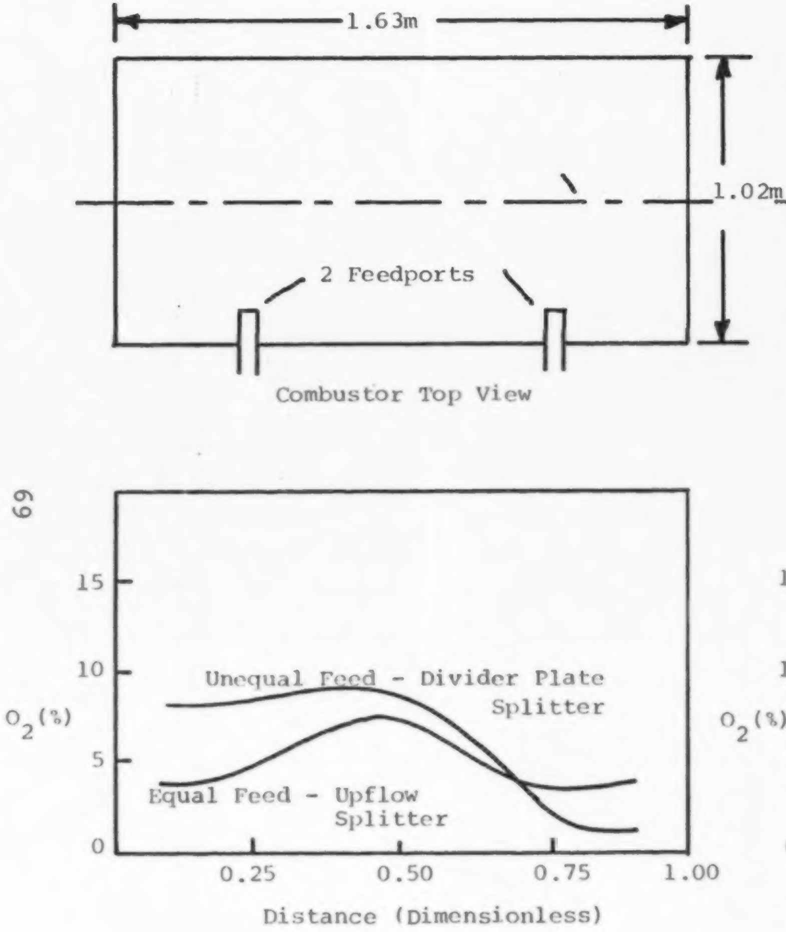


Figure 36. Freeboard survey - comparison upflow and divider plate feed splitters

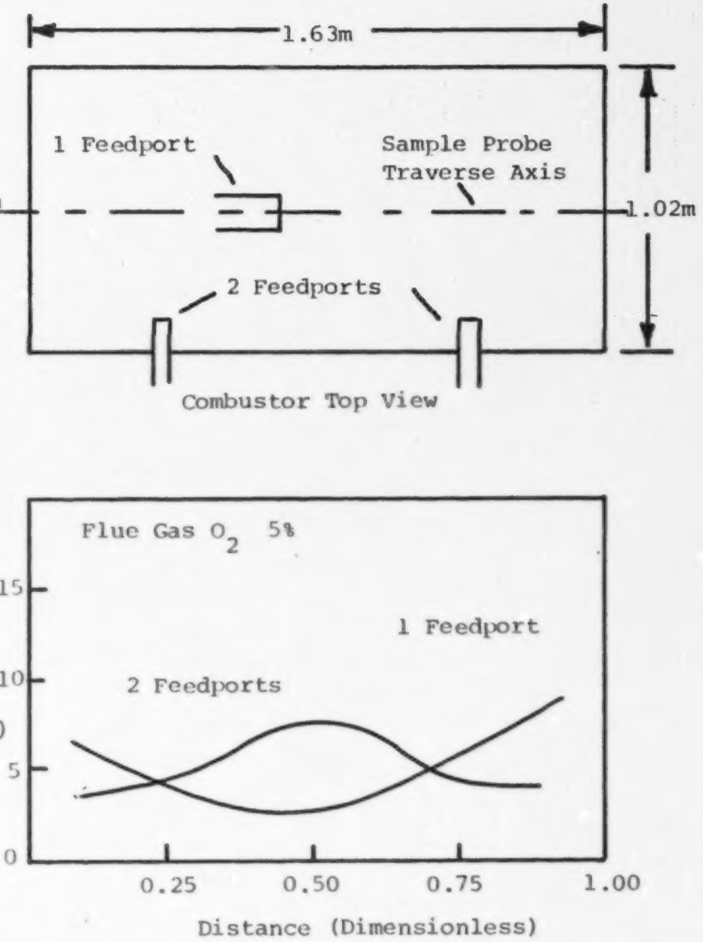


Figure 37. Freeboard survey - comparison one feed port vs. two feed ports

TABLE 1. TEST MATRIX TO ASSESS ONE VS. TWO FEED PORTS, AND FLY ASH RECYCLE VS. NO RECYCLE

Test No.	1	2	3	4	5	6	7	8	9	10	11	12
Sorbent Size, Mesh	-8	-8	-8	-8	-8	-8	4-30	4-30	4-30	4-30	4-30	4-30
Velocity m/sec	1.37	1.37	1.37	1.37	.91	.91	1.83	1.83	1.83	1.83	1.83	1.83
Bed Depth, m	0.91	0.91	0.91	0.91	1.37	1.37	1.37	1.37	0.91	0.91	0.91	0.91
Gas Residence, Time, Sec	.67	.67	.67	.67	1.5	1.5	.75	.75	.5	.5	.5	.5
Desired % SO_2 Removal	93	93	93	93	93	93	93	93	93	93	93	93
Fly Ash Recycle	No	Yes	Yes	No	No	No	No	No	No	No	Yes	Yes
No. of Feed Ports	1	1	2	2	2	1	1	2	1	2	2	1

Fuel - Illinois Coal

Sorbent - Greer Limestone

Bed Temp. - 843 C

Excess Oxygen - 5%

Sorbent feed rates were to be adjusted to achieve 93% SO_2 removal

TABLE 2. TEST MATRIX TO ASSESS EFFECTS OF GAS RESIDENCE TIME AND SORBENT PARTICLE SIZE

Test No.	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28
Sorbent Size, Mesh	4-30	4-30	4-30	4-30	4-30	4-30	4-30	4-30	-8	-8	-8	-8	-8	-8	-8	-8
Velocity, m/sec	1.83	1.83	1.83	1.37	1.37	1.83	1.83	1.83	.91	1.37	.91	.91	.91	.91	.91	1.37
Bed Depth, m	.91	.91	1.37	1.37	1.37	1.37	1.37	.91	.91	.91	.91	1.37	1.37	.91	.91	.91
Ca/S Molar Feed Ratio	2.0	2.75	2.75	2.75	2.0	2.0	2.75	2.75	2.0	2.75	2.75	2.75	2.0	2.0	2.75	2.0
Gas Residence Time, Sec.	.50	.50	.75	1.0	1.0	.75	.75	.75	1.0	.67	1.0	1.5	1.5	1.0	1.0	.67

Fuel - Illinois coal

Sorbent - Greer Limestone

Combustor Operation - Two feed ports and no fly ash recycle

Target SO_2 Removal - >90%

Bed Temp. - 843 C

Excess Oxygen - 5%

TABLE 3. SOLID SAMPLE ANALYSIS

Solid Stream	Frequency	Type of Analysis
Coal	Every 1-2 tests	Proximate, ultimate, size distribution
	Each test	Sulfur
Limestone	Two samples	Calcium, carbonate, magnesium, inerts
	Every 1-2 tests	Size distribution
Fly ash*	Each test	Calcium, sulfate, sulfur, carbonate, total carbon, size distribution, collection rate
Bed material	Each test	Calcium, sulfate, sulfur, carbonate, size distribution, bulk density, overflow rate

* Fly ash consists of the baghouse catch, and the cyclone catch when there is no recycle.

TABLE 4a. EMISSION TEST SUMMARY
(Tests 1 - 4)

Operating Conditions	FluiDyne Run No. (and Date) 24 (10/2) 25 (10/3)	Run 26 (10/4)	Run 27 (10/15)	Run 28 (10/16)	Run 28 (10/16)	Run 29 (10/18)
Test No.	1	2	3	3	3	4
Time interval at steady state, hrs	9	9	5	9	5.5	7
Bed temperature, °C	838	838	838	837	838	841
Superficial gas velocity, m/sec	1.34	1.34	1.35	1.32	1.35	1.35
Air flow rate, kg/min	42.6	42.3	42.7	41.9	42.5	42.5
Expanded bed depth, m	0.94	0.94	0.96	0.97	1.01	0.95
Gas residence time in bed, sec	0.70	0.70	0.71	0.73	0.75	0.70
Excess air, %	31.6	32.5	31.6	29.7	28.8	30.8
Coal feed rate, kg/hr	240	238	246	242	252	251
Limestone feed rate, kg/hr	86	76	78	61	75	66
Ca/S molar feed ratio[1], avg. (± std. deviation)	2.72	2.51 (0.12)	2.37 (0.13)	1.92 (0.20)	2.25 (0.06)	2.03 (0.11)
Number of feed ports	1	1	2	2	2	2
Fly ash recycle?	No	Yes	Yes	Yes	Yes	No
Flue gas emissions						
SO_2, ppm [1], avg. (± std. deviation)	161 (24)	195 (23)	163 (22)	191 (25)	222 (25)	221 (21)
CO_2, %	14.0	14.0	14.0	14.3	14.7	14.3
O_2, %	5.0	5.1	5.0	4.8	4.6	4.9

TABLE 4a. EMISSION TEST SUMMARY (cont'd)
(Tests 1 - 4)

		FluiDyne Run No. (and Date)				
Results	24 (10/2) 25 (10/3)	Run 26 (10/4)	Run 27 (10/15)	Run 28 (10/16)	Run 28 (10/16)	Run 29 (10/18)
Test No.	1	2	3	3	3	4
Sulfur Retention, %	94.2	92.8	94.4	93.4	92.5	92.4
Sulfation, %	34.6	37.0	39.8	49.0	41.1	45.5
lb SO_2/MMBtu	0.35	0.41	0.34	0.39	0.48	0.43
Combustion Eff. (% C in ash) [2]	96.1	97.6	-	97.5	97.5	97.0
Combustion Eff. (air burned), % [3]	97.1	95.7	95.4	94.6	94.6	94.4
Results Adjusted to 5% Excess Oxygen						
SO_2, ppm [4]	161	195	163	172	183	212
Sulfur retention, % [4]	94.2	92.8	94.4	94.0	93.8	92.7

---

(1) The 30-minute averages from the limestone feed and SO_2 traces are averaged to obtain the values reported in this table.

(2) Based upon the baghouse collection of elutriated carbon.

(3) Calculated using the air flow rate, excess air, coal feed rate and kg air/kg coal for stoichiometric combustion.

(4) Regression analysis used to determine effect of excess oxygen, and to adjust SO_2 emission level to a 5% oxygen baseline.

TABLE 4b. EMISSION TEST SUMMARY
(Tests 5 - 12)

	FluiDyne Run No. (and Date)							
Operating Conditions	Run 25 (1/21)	Run 37 (1/23)	Run 38 (1/30)	Run 39 (1/31)	Run 40 (2/5)	Run 41 (2/6)	Run 43 (2/8)	Run 44 (2/12)
Test No.	5	6	9	10	11	12	7	8
Time interval at steady state, hrs	6	5	8	3	6	6	5	4
Bed temperature, °C	840	840	845	840	840	840	840	837
Superficial gas velocity, m/sec	0.91	0.91	1.80	1.83	1.77	1.86	1.79	1.76
Air flow rate, kg/min	28.4	28.9	56.6	57.8	56.0	59.1	57.2	55.5
Expanded bed depth, m	1.27	1.32	1.01	0.97	0.99	0.96	1.37	1.35
Gas residence time in bed, sec	1.40	1.45	0.56	0.53	0.56	0.52	0.77	0.77
Excess air, %	29.2	30.4	31.6	31.6	29.2	41.5	30.8	31.6
Coal feed rate, kg/hr	172	162	322	320	330	303	327	339
Limestone feed rate, kg/hr	46	43	128	114	121	163	128	97
Ca/S molar feed ratio[1], avg. (± std. deviation)	2.04 (0.09)	1.90 (0.08)	2.85 (0.15)	2.50 (--)	2.47 (0.10)	3.76 (0.21)	2.69 (0.12)	2.02 (0.09)
Number of feed ports	2	1	1	2	2	1	1	2
Fly ash recycle?	No	No	No	No	Yes	Yes	No	No
Flue gas emissions								
SO_2, ppm [1], avg. (± std. deviation)	283 (33)	324 (33)	241 (44)	200 (--)	180 (24)	278 (54)	193 (18)	366 (13)
CO_2, %	-	14.5	14.0	--	14.0	--	14.5	--
O_2, %	4.7	4.8	5.1	5.0	4.7	6.1	4.9	5.0

TABLE 4b. EMISSION TEST SUMMARY (cont'd)
(Tests 5 - 12)

	FluiDyne Run No. (and Date)							
Results	Run 35 (1/21)	Run 37 (1/23)	Run 38 (1/30)	Run 39 (1/31)	Run 40 (2/5)	Run 41 (2/6)	Run 43 (2/8)	Run 44 (2/12)
Test No.	5	6	9	10	11	12	7	8
Sulfur Retention, %	90.7	89.1	92.0	93.0	94.6	90.0	93.8	88.8
Sulfation, %	44.5	46.9	32.3	37.2	38.3	23.9	34.9	44.0
lb SO_2/MMBtu	0.56	0.70	0.53	0.49	0.37	0.65	0.41	0.72
Combustion Eff. (% C in ash) (2)	97.1	98.0	97.1	--	98.0	98.5	96.2	96.0
Combustion Eff. (air burned), % (3)	92.0	97.0	94.8	--	90.7	95.1	91.7	91.8
Results Adjusted to 5% Excess Oxygen								
SO_2, ppm (4)	245	310	255	200	154	405	193	239
Sulfur retention, % (4)	91.8	89.5	91.5	93.0	95.3	87.1	93.8	88.8

---

(1) The 30-minute averages from the limestone feed and SO_2 traces are averaged to obtain the values reported in this table.

(2) Based upon the baghouse collection of elutriated carbon.

(3) Calculated using the air flow rate, excess air, coal feed rate and kg air/kg coal for stoichiometric combustion.

(4) Regression analysis used to determine effect of excess oxygen, and to adjust SO_2 emission level to a 5% oxygen baseline.

TABLE 4c. EMISSION TEST SUMMARY
(Tests 13 - 19)

	FluiDyne Run No. (and Date)						
Operating Conditions	Run 45 (4/29)	Run 45 (4/29)	Run 46 (4/30)	Run 46 (4/30)	Run 47 (5/2)	Run 48 (5/5)	Run 48 (5/5) Run 49 (5/6)
Test No.	13	14	15	16	17	18	19
Time interval at steady state, hrs	7	4.5	3	3	2.5	1.5	2
Bed temperature, °C	837	837	842	842	843	840	842
Superficial gas velocity, m/sec	1.74	1.77	1.79	1.40	1.38	1.77	1.83
Air flow rate, kg/min	54.9	55.6	55.8	44.0	43.3	55.5	57.7
Expanded bed depth, m	1.03	0.92	1.38	1.36	1.28	1.34	1.37
Gas residence time in bed, sec	0.59	0.52	0.77	0.98	0.93	0.76	0.75
Excess air, %	31.2	25.1	32.7	30.7	32.7	33.8	30.3
Coal feed rate, kg/hr	316	350	332	267	260	324	343
Limestone feed rate, kg/hr	102	147	133	105	70	99	140
Ca/S molar feed ratio(1), avg. (± std. deviation)	2.20 (0.20)	2.89 (--)	2.91 (--)	2.88	2.00	2.02	2.95
Number of feed ports	2	2	2	2	2	2	2
Fly ash recycle?	No	No	No	No	No	No	No
Flue gas emissions							
SO_2, ppm (1)	305	339	80	37	203	472	289
O_2, %	5.0	4.2	5.1	4.0	5.1	5.2	4.8

TABLE 4c. EMISSION TEST SUMMARY (cont'd)
(Tests 13 - 19)

	FluiDyne Run No. (and Date)						
Results	Run 45 (4/29)	Run 45 (4/29)	Run 46 (4/30)	Run 46 (4/30)	Run 47 (5/2)	Run 48 (5/5)	Run 48 (5/5) Run 49 (5/6)
Test No.	13	14	15	16	17	18	19
Sulfur Retention, %	90.5	90.4	97.4	98.8	93.4	85.9	90.8
Sulfation, %	41.1	31.3	33.5	34.3	46.7	42.5	30.8
lb SO_2/MMBtu							
Combustion Eff. (% C in ash) (2)	96.0	97.3	95.6	96.8	96.4	94.8	94.1
Combustion Eff. (air burned), % (3)	91.4	87.5	90.6	91.9	89.9	91.4	92.4
Results Adjusted to 5% Excess Oxygen							
SO_2, ppm (4)	297	204	85	35	215	529	259
Sulfur retention, % (4)	90.7	93.9	97.3	98.9	93.1	84.5	91.4

---

(1) The 30-minute averages from the limestone feed and SO_2 traces are averaged to obtain the values reported in this table.

(2) Based upon the baghouse collection of elutriated carbon.

(3) Calculated using the air flow rate, excess air, coal feed rate and kg air/kg coal for stoichiometric combustion.

(4) Regression analysis used to determine effect of excess oxygen, and to adjust SO_2 emission level to a 5% oxygen baseline.

TABLE 4d. EMISSION TEST SUMMARY
(Tests 20 - 26)

	FluiDyne Run No. (and Date)						
Operating Conditions	Run 50 (5/7)	51 (5/8) 52 (5/9)	52 (5/9) 53 (5/12)	Run 53 (5/12) Run 54 (5/13)	Run 54 (5/13)	Run 55 (5/14)	Run 56 (5/15)
Test No.	20	21	22	23	24	25	26
Time interval at steady state, hrs	20	25	3	4	6	8	8
Bed temperature, °C	836	839	845	848	849	843	842
Superficial gas velocity, m/sec	1.83	0.94	1.28	0.98	0.95	0.96	0.92
Air flow rate, kg/min	57.7	29.7	40.2	30.5	29.6	30.2	29.1
Expanded bed depth, m	0.99	0.95	1.05	0.91	1.29	1.29	0.99
Gas residence time in bed, sec	0.54	1.00	0.82	0.94	1.37	1.35	1.08
Excess air, %	34.7	33.0	32.2	30.3	31.3	30.4	35.3
Coal feed rate, kg/hr	318	166	235	181	174	166	155
Limestone feed rate, kg/hr	119	48	90	68	61	46	45
Ca/S molar feed ratio$^{(1)}$	2.61	2.02	2.68	2.65	2.36	1.95	1.95
Number of feed ports	2	2	2	2	2	2	2
Fly ash recycle?	No	No	No	No	No	No	No
Flue gas emissions							
SO_2, ppm $^{(1)}$	269	258	207	171	119	155	304
O_2, %	5.4	5.2	5.1	4.8	5.0	4.8	5.4

TABLE 4d. EMISSION TEST SUMMARY (cont'd)
(Tests 20 - 26)

Results	Run 50 (5/7)	51 (5/8) 52 (5/9)	52 (5/9) 53 (5/12)	Run 53 (5/12) Run 54 (5/13)	Run 54 (5/13)	Run 55 (5/14)	Run 56 (5/15)
	FluiDyne Run No. (and Date)						
Test No.	20	21	22	23	24	25	26
Sulfur Retention, %	91.0	91.4	93.5	94.6	96.4	94.8	89.9
Sulfation, %	34.9	45.2	34.9	35.6	40.8	48.6	46.1
lb SO_2/MMBtu							
Combustion Eff. (% C in ash) (2)	96.5	97.3	97.6	98.2	97.5	98.3	98.0
Combustion Eff. (air burned), % (3)	96.1	96.1	91.7	91.0	89.1	95.6	94.6
Results Adjusted to 5% Excess Oxygen							
SO_2, ppm (4)	306	274	214	160	118	144	354
Sulfur retention, % (4)	90.0	91.0	93.4	94.9	96.4	95.1	88.6

---

(1) The 30-minute averages from the limestone feed and SO_2 traces are averaged to obtain the values reported in this table.

(2) Based upon the baghouse collection of elutriated carbon.

(3) Calculated using the air flow rate, excess air, coal feed rate and kg air/kg coal for stoichiometric combustion.

(4) Regression analysis used to determine effect of excess oxygen, and to adjust SO_2 emission level to a 5% oxygen baseline.

TABLE 4e. EMISSION TEST SUMMARY
(Tests 27 - 28)

Operating Conditions	FluiDyne Run No. (and Date)	
	Run 56 (5/15) Run 57 (5/16)	Run 57 (5/16)
Test No.	27	28
Time interval at steady state, hrs	5	3
Bed temperature, °C	849	846
Superficial gas velocity, m/sec	0.98	1.31
Air flow rate, kg/min	30.6	41.0
Expanded bed depth, m	1.00	0.99
Gas residence time in bed, sec	1.03	0.75
Excess air, %	31.3	32.4
Coal feed rate, kg/hr	166	224
Limestone feed rate, kg/hr	64	67
Ca/S molar feed ratio(1)	2.57	1.90
Number of feed ports	2	2
Fly ash recycle?	No	No
Flue gas emissions		
SO_2, ppm (1)	143	301
O_2, %	5.0	5.1

TABLE 4e. EMISSION TEST SUMMARY (cont'd)
(Tests 27 - 28)

	FluiDyne Run No. (and Date)	
Results	Run 56 (5/15) Run 57 (5/16)	Run 57 (5/16)
Test No.	27	28
Sulfur Retention, %	95.3	90.8
Sulfation, %	37.1	47.8
lb SO_2/MMBtu		
Combustion Eff. (% C in ash) (2)	98.2	98.1
Combustion Eff. (air burned), % (3)	95.9	94.2
Results Adjusted to 5% Excess Oxygen		
SO_2, ppm (4)	135	314
Sulfur retention, % (4)	95.6	90.4

---

(1) The 30-minute averages from the limestone feed and SO_2 traces are averaged to obtain the values reported in this table.

(2) Based upon the baghouse collection of elutriated carbon.

(3) Calculated using the air flow rate, excess air, coal feed rate and kg air/kg coal for stoichiometric combustion.

(4) Regression analysis used to determine effect of excess oxygen, and to adjust SO_2 emission level to a 5% oxygen baseline.

TABLE 5. RESPONSE OF SO_2 EMISSIONS TO SHIFTS IN THE FLUE GAS EXCESS OXYGEN LEVEL

Superficial gas velocity (m/sec)	Average Response (ppm change in SO_2 per 0.1% increase in O_2)
0.9	- 8.5
1.4	- 9.4
1.8	-12.9

Notes:

1. The response figures shown (ppm SO_2/0.1% O_2) are the averages obtained for all tests at the indicated superficial gas velocity.

2. These responses apply for O_2 levels in the range tested here (i.e., levels of about 5% O_2, or 30% excess air).

3. Negative values for the responses indicate that SO_2 levels in the flue gas decrease as O_2 levels increase.

4. A 10 ppm decrease in flue gas SO_2 concentration corresponds to approximately a 0.3 percentage point increase in SO_2 retention.

TABLE 6. RESPONSE OF SULFUR RETENTION TO THE TEST VARIABLES

Variable	Range	Sulfur Retention (% Increase)
Ca/S Ratio	1.9 → 2.95	6
Excess Oxygen (%)	5.0 → 6.1	3
Limestone Feed Size (Mesh) (Mass mean, microns)	4x30 → -8 1300 → 800	2
Gas/Residence Time (Sec)	0.5 → 1.5	2
Fly Ash Recycle	No → Yes	2
Feed Ports	1 → 2	2
Regression analysis - standard error of estimate		± 2.4
Probable measurement error at 95% confidence level		± 0.9

TABLE 7. RESIDUALS ANALYSIS OF THE EXPERIMENTAL SCATTER REMAINING IN THE SULFUR RETENTION DATA AFTER REGRESSION ANALYSIS

Changes in Operating Conditions Resulting in Steady-State Being Approached from the High Side	Average Residual ($R_{Meas.} - R_{Pred.}$)
Ca/S Ratio Down	+1.2
Velocity Down	+1.9
Bed Level Up	+0.7
Changes in Operating Conditions Resulting in Steady-State Being Approached from the Low Side	
Ca/S Ratio Up	-1.5
Velocity Up	-3.6
Bed Level Down	-3.0
No Change in Operating Conditions	+0.75

TABLE 8. REPLICATE TEST CONDITIONS PERFORMED DURING TESTS 13-28

Replicate Test No's	Ca/S Ratio (Mole/Mole)	Residence Time (Sec)	Sulfur Retention (%)	Difference (%)
21	2.02	1.0	91.0	2.4
26	1.95	1.0	88.6	
23	2.65	.94	94.9	0.7
27	2.57	1.03	95.6	
15	2.9	.77	97.3	5.9
19	2.95	.75	91.4	
14	2.9	.52	93.9	3.9
20	2.6	.54	90.0	

TABLE 9. COMPOSITION (WT%) OF SORBENTS SCREENED FOR THE TEST PROGRAM

Supplier	Location	$CaCO_3$	$MgCO_3$	Inerts
J.L. Shiely Co.	St. Paul, MN	51.5	36.6	11.9
B.L. Anderson Co.	Cedar Rapids, IA			
Alpha Industries	Marion, IA			
Hactor Construction Co.	Caledonia, MN	43.5	45.5	11.0
Lundeen Quarry (1)	Owatonna, MN	54.0	17.7	33.0
Lundeen Quarry (2)	Owatonna, MN	43.6	29.4	27.0
E.E. Duff & Sons	Huntsville, OH	51.8	43.3	4.9
Calcium Carbonate Co.	Quincy, IL	97.0	1.7	1.3
B.L. Anderson Co.	Cedar Rapids, IA	88.5	3.4	8.1
Walp McCartent Co.	Fort Dodge, IA	93.6	4.8	1.6
Greer Limestone Co.	Morgantown, W. VA	78.5	3.2	18.3
Grove Limestone Co.	Stephen City, VA	94.1	1.0	4.9

---

(1) Quarry deposits variable - composition of old shipment

(2) Quarry deposits variable - composition of new shipment

Analysis performed by Westinghouse

TABLE 10. WESTINGHOUSE SORBENT SCREENING TESTS -- REACTIVITY AND ATTRITION RESISTANCE (Ref. 4)

Sorbent	Attrition (1)	Reactivity (2)
	Difference Between Percent of Particles < 32 Mesh in Samples After and Before Attrition	mg SO_3/mg Stone
Group 1		
Lundeen Dolomite (Old Shipment)	-0.3	0.31
Anderson Dolomite	5.9	
Shiely Dolomite	7.6	0.22
Alpha Dolomite	12.7	
Group 2		
Hactor Dolomite	12.9	0.25
Anderson Limestone	5.9	0.30
McCarten Limestone	6.5	0.26
Ref. Sorbents (3)		
Greer Limestone	0.7	
Duff Dolomite	0.4	
Grove Limestone	0.14	
Calcium Carbonate Co. Limestone (4)	31.7	0.25

---

(1) Performed in a bench scale fluidized bed with a calcined stone.

(2) Sorbent capacity determined by TGA tests.

(3) Sorbents which have a record of successful use in the FBC community. Attrition tests were performed under a variety of conditions and in different equipment in some cases.

(4) Test to diagnose operating problems with this sorbent.

TABLE 11. ELUTRIATION OF CALCIUM CARBONATE CO. LIMESTONE

Time Period	Coal Feed Rate (Kg/hr)	Lime-stone Feed Rate (Kg/hr)	Lime-stone Size (Mesh)	Bag-house Collec-tion (Kg/hr)	Feed Limestone Elutriated as % of Feed	Bed Limestone Elutriated as % of Feed	Total Limestone Elutriated as % of Feed	Bed Level Change Calcu-lated (cm/hr)	Bed Level Change Measured (cm/hr)
13:40 15:40	168	0	--	43	--	--	--	-1.5	-1
15:55 17:55	164	59	20x60	70	62	50	112	-.4	~0
21:35 22:40	156	79	20x60	80	66	41	107	-.1	~0
23:05 00:05	161	79	6x20	73	56	40	96	+.3	~0

TABLE 12. WESTINGHOUSE ATTRITION TESTS PERFORMED TO EVALUATE THE OPERATING PROBLEMS WITH CALCIUM CARBONATE CO. LIMESTONE AND LUNDEEN DOLOMITE

	Attrition(1)	
Sorbent	Difference Between Percent of Particles < 32 Mesh in Samples After and Before Attrition	Comments
Calcium Carbonate Co. Limestone	31.7	
Lundeen Dolomite (Old shipment)	-0.3	Sample taken from remaining part of shipment used in previous test program.
Lundeen Dolomite (New shipment - dried)	0.33	Sample taken from shipment received for test program. Material dried at agrigate plant with rotary drier.
Lundeen Dolomite (New shipment - not dried)	3.3	Sample direct from quarry without drying.
Lundeen Dolomite (New shipment - uncalcined)		Attrition test performed with unheated fluidized bed to evaluate the material as-received.
Candidate Sorbents range of values except for Lundeen Dolomite	5.9 to 12.9	Sorbents listed as Group 1 and Group 2 in Table 10.
Reference Sorbents	0.14 to 0.7	Sorbents with a record of successful use in FBCs. Table 10.

---

(1) Performed in a bench scale fluidized bed with a calcined stone.

TABLE 13. VARIABLE COMPOSITION OF THE DOLOMITE SHIPMENTS RECEIVED FROM THE LUNDEEN QUARRY

	$CaCO_3$ (Wt %)	$MgCO_3$ (Wt %)	Inerts (Wt %)
Lundeen Dolomite (1) (Old Shipment)	54.0	17.7	33.0
Lundeen Dolomite (2) (New Shipment)	43.6	29.4	27.0

---

(1) Sample from remaining part of shipment received for a previous test program.

(2) Sample from shipment received for the EPA test program.

TABLE 14. LOS ANGELES ABRASION TESTS

Type of Material	Attrition Loss (%)
Greer Limestone	11.7
Lundeen Dolomite (Old shipment)	20.6
Lundeen Dolomite (New shipment)	28.6

Test Method - 5000 gm 8 to 20 mesh sample rescreened on 20 mesh, 5 steel spheres, 350 revolutions

TABLE 15. UPFLOW FEED DIVIDER TESTS

Test No.	Total Feed Rate (kg/hr)	Deviation of Split From Average (%)
1	79	± 3.5
2	157	± 2.1
3	213	± .5
4	233	± .3

Transport Air Flow rate 250 kg/hr

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APPENDIX A. RUN SUMMARY

Run No.	Activities
1 - 4	-Shakedown tests
5 - 10	-Tests with Calcium Carbonate Co. limestone
11 - 15	-Tests with Lundeen dolomite
16 - 22	-Greer limestone bed sulfated
	-Freeboard emissions profile surveyed to evaluate feed port location and the feed divider; "upflow" divider installed.
	-Freeboard burning identified - spray nozzle installed to cool flue gas.
23 - 31	-Test 1-4 completed
	-Several runs unproductive because of high elutriation rates with a deep bed; roof level raised.
32 - 44	-Tests 5-12
45 - 57	-Tests 13-28

APPENDIX B. COMPARISON OF SO_2 MEASUREMENTS BY DUPONT ULTRAVIOLET AND WET CHEMISTRY METHODS

EPA Test No.	FluiDyne Test No.	SO_2 (PPM)		% Difference
		UV	WC	$\frac{UV - WC}{UV} \times 100$
-	13	1237	1114	+10
1	24	197	202	- 3
2	26	165	167	- 1
3	28	195	213	- 9
4	29	120	148	-23
5	35	267	296	-11
6	37	328	316	+ 4
7	43	180	212	-17
8	44	170	183	- 8
9	38	175	179	- 2
11	40	130	121	+ 7
12	41	340	418	-23
13	45	417	281	+33
15	46	84	58	+31
17	47	86	77	+10
19	49	306	291	+ 5
20	50	299	251	+16
22	53	250	240	+ 4
24	54	180	152	+16

APPENDIX B. COMPARISON OF SO_2 MEASUREMENTS BY DUPONT ULTRAVIOLET AND WET CHEMISTRY METHODS (CONT'D)

EPA Test No.	FluiDyne Test No.	SO_2 (PPM) UV	SO_2 (PPM) WC	% Difference $\frac{UV - WC}{UV} \times 100$
25	55	132	131	+ 1
26	56	291	222	+24
27	57	110	80	+27

- -

Average	Test 1 - 12	-6 ± 11
Average	Test 13 - 28	+17 ± 12 (1)
Average	Test 1 - 28	+4 ± 16

---

(1) Wet chemistry analysis performed by different technician

APPENDIX C. FLUE GAS NO_X MEASUREMENTS BY UV AND WET CHEMISTRY ANALYSIS (EPA METHOD 7)

EPA Test No.	FluiDyne Run No.	NO_X (PPM) UV	NO_X (PPM) WC	O2 %	LB NO_X/ MMBTU	% Difference $\frac{UV - WC}{UV} \times 100$
1	24	308	324	5.7	0.47	+ 3
		360		5.7	0.55	
1	25	300	--	5.2	0.46	--
		330		5.5	0.50	
2	26	260	241	5.0	0.40	+ 5
		270		5.0	0.42	
3	28	275	250	4.8	0.41	+13
		300		4.8	0.45	
4	29	210	221	4.5	0.31	+11
		285		4.5	0.42	
5	35	190	--	4.6	0.27	--
		210		4.6	0.30	
6	37	190	200	5.0	0.37	0
		215		4.6	0.39	
7	43	240	254	5.1	0.35	- 6
8	44	265	258	5.0	0.39	+ 8
		295		5.0	0.44	
10	39	250	330	5.0	0.37	-28
		265		5.3	0.39	
11	40	220	331	4.4	0.32	-48
		225		4.8	0.32	
12	41	330	288	6.5	0.54	+14
		340		6.5	0.56	
15	46	460	--	5.1	0.69	--
		460		5.6	0.69	

APPENDIX C. FLUE GAS NO_x MEASUREMENTS BY UV AND WET CHEMISTRY ANALYSIS (EPA METHOD 7) (CONT'D)

EPA Test No.	FluiDyne Run No.	NO_x (PPM) UV	NO_x (PPM) WC	02 %	Lb NO_x/ MMBTU	% Difference $\frac{UV - WC}{UV}$ x 100
17	47	275 342	--	4.0 5.2	0.44 0.54	--
21	52	305 310	240	4.8 4.9	0.49 0.49	+22
22	53	225 330	192	5.2 6.0	0.34 0.51	+30
24	54	245 260	210	4.6 5.0	0.36 0.38	+17
25	55	285 295	239	4.8 4.8	0.43 0.45	+18
26	56	270 280	--	5.2 5.4	0.42 0.44	--
27	57	320 325 350	295	4.7 4.8 5.0	0.51 0.52 0.56	+11
	Average	Tests 1 - 12			6 ± 7 (1)	
		Tests 21 - 27			20 ± 7 (2)	
		Overall			11 ± 10	

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(1) Two points excluded + -48, -28

(2) Wet chemistry analysis performed by different technician

APPENDIX D. ENTRAINMENT LOSSES WITH GREER LIMESTONE

Run No.	Ca/S Feed Ratio (Mole/Mole)	Entrainment Losses Ca/S Equiv. (Mole/Mole)	Wt. Percent Input Limestone Lost by Entrainment
25 (Test 1)	2.72	1.06	39
26 (Test 2)	2.51	.85	34
28 (Test 3)	1.92	.69	36
29 (Test 4)	2.03	1.14	56
35 (Test 5)	1.96	.92	47
37 (Test 6)	1.88	.86	46
38 (Test 9)	2.54	.86	34
40 (Test 11)	2.44	.56	23
41 (Test 12)	3.6	1.19	33
43 (Test 7)	2.76	2.62	95
44 (Test 8)	2.02	2.48	123
45 (Test 13)	2.20	1.36	62
45 (Test 14)	2.89	.78	27
46 (Test 15)	2.91	3.20	110
46 (Test 16)	2.88	2.88	100
47 (Test 17)	2.0	1.80	90
48 (Test 18)	2.02	1.98	98
49 (Test 19)	2.95	3.27	111
50 (Test 20)	2.61	.84	32
52 (Test 21)	2.02	.57	28
53 (Test 22)	2.68	1.07	40
53 (Test 23)	2.66	1.30	49
54 (Test 24)	2.36	1.82	77
55 (Test 25)	1.95	1.21	62
56 (Test 26)	1.95	.66	34
57 (Test 27)	2.57	.85	33
57 (Test 28)	1.90	.61	32

APPENDIX E. BAGHOUSE PARTICULATE EMISSION

FluiDyne Run No.	Baghouse Particulate Loading			Isokinetic %	Baghouse Air/Cloth Ratio (ACFM/ft^2)	Particulate Emission LB/M BTU
	Lb/Hr in	Lb/Hr out	Grains/SCF in			
11	150	.0027	10.6	101	2.4	.001
18	94	.24	4.1	109	4.2	.041
24 (Test 1)	175	.174	7.3	104	4.2	.032
26 (Test 2)	108	.065	4.5	101	4.3	.011
28 (Test 3)	115	.11	4.5	108	4.7	.019
29 (Test 4)	133	.136	6.4	105	3.6	.024
33	57	.0038	4.6	99	2.1	.001
34	86	.0113	5.6	104	2.6	.003
35 (Test 5)	91	.068	8.5	105	1.9	.017
37 (Test 6)	78.8	.021	9.0	112	1.5	.005
40 (Test 11)	134	.046	4.8	114	5.1	.031
41 (Test 12)	134	.126	5.8	129	4.6	.017
43 (Test 7)	294	1.03	10.4	103	5.5	.13

APPENDIX E. BAGHOUSE PARTICULATE EMISSION (CONT'D)

FluiDyne Run No.	Baghouse Particulate Loading			Isokinetic %	Baghouse Air/Cloth Ratio (ACFM/ft^2)	Particulate Emission LB/M BTU
	Lb/Hr in	Lb/Hr out	Grains/SCF in			
44 (Test 8)	307	2.43	11.2	102	5.4	.32
45 (Test 13)	198	8.89	7.7	100	4.9	1.20
45 (Test 13)	198	9.62	7.4	98	5.1	1.28
46 (Test 15)	329	6.78	11.26	90	5.6	1.11
47 (Test 17)	170	.294	8.8	98.1	3.6	.049
48 (Test 18)	282	.395	11.0	93.4	4.9	.054
49 (Test 19)	387	.098	14.7	99	5.2	.013
50 (Test 20)	146	.095	5.4	104	5.1	.013
52 (Test 21)	70	.024	8.7	96	1.4	.006
53 (Test 22)	105	.020	6.3	103	3.2	.004
54 (Test 23)	107	.032	8.2	104	2.3	.007

APPENDIX E. BAGHOUSE PARTICULATE EMISSION (CONT'D)

FluiDyne Run No.	Baghouse Particulate Loading			Isokinetic %	Baghouse Air/Cloth Ratio (ACFM/ft^2)	Particulate Emission LB/M BTU
	Lb/Hr in	Lb/Hr out	Grains/SCF in			
54 (Test 24)	110	.015	9.1	102	2.3	.003
55 (Test 25)	95	.017	8.6	103	2.0	.004
56 (Test 26)	67	.028	5.5	100	2.2	.007

- Isokinetics average 103 ± 7
- Emissions were high during runs 43-46 because flue gas bypassed the bags when the bag holding flanges loosened.

APPENDIX F. IMPACTOR SIZE DISTRIBUTION DATA ACQUIRED UPSTREAM OF THE BAGHOUSE

	Particulate Mass Flow						Results				
FluiDyne Run No.	Baghouse Collect. (PPH)	Impactor Mass Flow (PPH)	Grains/ SCF	Isokinetic (%)	Impact Flow ACFM		Probe	Stage 1	Stage 2	Stage 3	Stage 4
8	-	243	25	102	0.69	Weight collect (gr)		21.9	.19	.06	.01
						Weight percent < Stage		1.16	.33	.06	
						Cut Point (µm)		10.2	5.6	1.6	
20	-	95	6.3	100	1.09	Weight collect (gr)	.71	12.8	2.36	2.0	.43
						Weight percent < Stage		26.1	17.1	2.3	
						Cut Point (µm)		8.8	4.5	1.2	

APPENDIX F. IMPACTOR SIZE DISTRIBUTION DATA ACQUIRED UPSTREAM OF THE BAGHOUSE (CONT'D)

	Particulate Mass Flow							Results			
FluiDyne Run No.	Baghouse Collect. (PPH)	Impactor Mass Flow (PPH)	Grains/ SCF	Isokinetic (%)	Impact Flow ACFM		Probe	Stage 1	Stage 2	Stage 3	Stage 4
25 (Test 1)	126	141	9.5	102	1.14	Weight collect (gr)	1.27	18.4	5.98	1.68	.06
						Weight percent < Stage		28.2	6.3	.2	
						Cut Point (µm)		8.3	4.7	1.2	
28 (Test 3)	115	84	5.3	139	.68	Weight collect (gr)	.82	6.83	1.07	.85	.12
						Weight percent < Stage		21.1	10.0	1.2	
						Cut Point (µm)		10.6	5.6	1.5	

APPENDIX F. IMPACTOR SIZE DISTRIBUTION DATA ACQUIRED UPSTREAM OF THE BAGHOUSE (CONT'D)

	Particulate Mass Flow						Results				
FluiDyne Run No.	Baghouse Collect. (PPH)	Impactor Mass Flow (PPH)	Grains/ SCF	Isokinetic (%)	Impact Flow ACFM		Probe	Stage 1	Stage 2	Stage 3	Stage 4
32		32	2.9	101	.37	Weight collect (gr)		1.97	.31	.47	.09
						Weight percent < Stage		30.7	19.7	3.0	
						Cut Point (μm)		14.0	7.7	2.3	
33	57	86	7.5	99	.30	Weight collect (gr)	1.28	3.58	.71	1.17	.52
						Weight percent < Stage		33.0	23.2	7.1	
						Cut Point (μm)		14.6	8.5	2.7	

APPENDIX F. IMPACTOR SIZE DISTRIBUTION DATA
ACQUIRED UPSTREAM OF THE BAGHOUSE (CONT'D)

	Particulate Mass Flow						Results				
FluiDyne Run No.	Baghouse Collect. (PPH)	Impactor Mass Flow (PPH)	Grains/ SCF	Isokinetic (%)	Impact Flow ACFM		Probe	Stage 1	Stage 2	Stage 3	Stage 4
34	86	100	9.3	108	0.82	Weight collect (gr)	2.35	11.9	2.33	2.62	.46
						Weight percent < Stage		27.5	15.7	2.4	
						Cut Point (µm)		9.6	5.1	1.4	
35 (Test 5)	91	78	8.6	107	.71	Weight collect (gr)	1.43	9.52	1.47	2.46	.53
						Weight percent < Stage		28.9	19.4	3.4	
						Cut Point (µm)		10.4	5.4	1.5	

APPENDIX F. IMPACTOR SIZE DISTRIBUTION DATA ACQUIRED UPSTREAM OF THE BAGHOUSE (CONT'D)

	Particulate Mass Flow						Results				
FluiDyne Run No.	Baghouse Collect. (PPH)	Impactor Mass Flow (PPH)	Grains/ SCF	Isokinetic (%)	Impact Flow ACFM		Probe	Stage 1	Stage 2	Stage 3	Stage 4
37 (Test 6)	81	59	6.7	121	.95	Weight collect (gr)	1.89	7.8	1.37	2.07	.17
						Weight percent < Stage		27.1	16.8	1.3	
						Cut Point (μm)		9.0	4.9	1.3	
40 (Test 11)	134	179	14.1	132	.38	Weight collect (gr)	.63	6.72	1.32	.85	.27
						Weight percent < Stage		24.9	11.4	2.7	
						Cut point (μm)		14.0	7.5	2.2	

APPENDIX F. IMPACTOR SIZE DISTRIBUTION DATA
ACQUIRED UPSTREAM OF THE BAGHOUSE (CONT'D)

	Particulate Mass Flow							Results			
FluiDyne Run No.	Baghouse Collect. (PPH)	Impactor Mass Flow (PPH)	Grains/ SCF	Isokinetic (%)	Impact Flow ACFM		Probe	Stage 1	Stage 2	Stage 3	Stage 4
41 (Test 12)	134	82	6.5	116	.38	Weight collect (gr)	1.11	2.55	.5	.71	.17
	Probe heater burned out condensation in line					Weight percent <Stage		27.4	17.5	3.3	
						Cut Point (µm)		14.0	7.5	2.2	
43 (Test 7)	294	149	8.2	104	.37	Weight collect (gr)	1.14	5.38	.47	.27	.33
						Weight percent < Stage		14.0	7.8	4.3	
						Cut Point (µm)		14.0	7.5	2.2	

APPENDIX F. IMPACTOR SIZE DISTRIBUTION DATA ACQUIRED UPSTREAM OF THE BAGHOUSE (CONT'D)

	Particulate Mass Flow						Results				
FluiDyne Run No.	Baghouse Collect. (PPH)	Impactor Mass Flow (PPH)	Grains/ SCF	Isokinetic (%)	Impact Flow ACFM		Probe	Stage 1	Stage 2	Stage 3	Stage 4
44	304	151	8.8	104	.33	Weight collect (gr)	5.69	.63	.32	.68	.19
		Probe heater burned out condensation in line				Weight percent < Stage		15.8	11.6	2.6	
						Cut Point (μm)		14.3	8.2	2.5	
45 (Test 13)	198	98	7.3	107	.39	Weight collect (gr)	3.7	.40	.36	.67	.08
		Probe contained much material				Weight percent < Stage		20.2	13.1	1.5	
						Cut Point (μm)		13.8	7.3	2.2	

APPENDIX F. IMPACTOR SIZE DISTRIBUTION DATA ACQUIRED UPSTREAM OF THE BAGHOUSE (CONT'D)

	Particulate Mass Flow						Results				
FluiDyne Run No.	Baghouse Collect. (PPH)	Impactor Mass Flow (PPH)	Grains/ SCF	Isokinetic (%)	Impact Flow ACFM		Probe	Stage 1	Stage 2	Stage 3	Stage 4
46 (Test 15)	329	110	6.9	118	.39	Weight collect (gr)	3.5	2.2	.14	.34	107
		Probe plugged				Weight percent <Stage		8.8	6.5	1.1	
						Cut Point (µm)		13.8	7.3	2.2	
47	170	137	10.2	138	.63	Weight collect (gr)		.98	8.4	1.18	.24
						Weight percent < Stage		19.3	12.3	2.0	
						Cut Point (µm)		10.9	5.8	1.7	

APPENDIX F. IMPACTOR SIZE DISTRIBUTION DATA
ACQUIRED UPSTREAM OF THE BAGHOUSE (CONT'D)

	Particulate Mass Flow						Results				
FluiDyne Run No.	Baghouse Collect. (PPH)	Impactor Mass Flow (PPH)	Grains/ SCF	Isokinetic (%)	Impact Flow ACFM		Probe	Stage 1	Stage 2	Stage 3	Stage 4
52	70	50	5.2	99	0.47	Weight collect (gr)	1.07	3.12	.56	.66	.97
						Weight percent < Stage		34.3	25.5	15.2	
						Cut Point (m)		12.8	6.8	2.0	
53	105	119	8.8	109	.51	Weight collect (gr)	1.66	6.68	.84	1.30	.26
						Weight percent < Stage		22.3	14.5	2.4	
						Cut Point (μm)		12.2	6.5	1.9	
55 (Test 25)	95	80	8.3	100	.41	Weight collect (gr)	.97	5.98	.52	.89	.14

APPENDIX F. IMPACTOR SIZE DISTRIBUTION DATA ACQUIRED UPSTREAM OF THE BAGHOUSE (CONT'D)

	Particulate Mass Flow							Results			
FluiDyne Run No.	Baghouse Collect. (PPH)	Impactor Mass Flow (PPH)	Grains/ SCF	Isokinetic (%)	Impact Flow ACFM		Probe	Stage 1	Stage 2	Stage 3	Stage 4
55 (con't)						Weight percent < Stage		18.2	12.1	1.6	
						Cut Point (μm)		13.8	7.3	2.2	
56 (Test 26)	67	67	7.3	106	.41	Weight collect (gr)	1.11	4.56	.58	1.04	.17
						Weight percent < Stage		24.0	16.2	2.3	
						Cut Point (μm)		13.8	7.3	2.2	

APPENDIX G. ILLINOIS COAL COMPOSITION ULTIMATE ANALYSIS

FluiDyne Run No.	EPA Test No.	Moisture	Ash	Total Carbon	H	S	N	Cl	O_2	Heating Value BTU/Lb
26	2	10.2	12.1	61.2	4.3	3.23	1.1	.07	7.69	11104
29	4	9.9	12.2	61.5	4.2	3.27	1.1	.04	7.75	11087
35	5	11.6	11.6	60.7	4.3	3.21	1.1	.07	7.4	10970
37	6	8.7	12.4	62.0	4.4	3.48	1.1	.05	7.8	11186
40	11	5.3	13.0	64.1	4.6	3.67	1.1	.06	8.2	11616
43	7	6.1	11.8	65.5	4.6	3.71	1.1	.06	7.3	11745
45	14	6.1	12.3	64.0	4.5	3.65	1.2	.06	8.4	11573
46	16	6.1	14.9	61.7	4.3	3.44	1.2	.06	8.3	11076
48	18	5.9	15.7	61.3	4.4	3.79	1.2	.05	7.7	11034
50	20	6.1	14.6	61.8	4.3	3.60	1.1	.05	8.5	11063
53	23	6.4	13.0	62.8	4.3	3.62	1.2	.04	8.7	11345
54	24	5.5	12.1	64.4	4.5	3.74	1.2	.05	8.5	11658
56	26	6.9	11.3	64.4	4.6	3.75	1.1	.06	7.9	11624

Analysis made on composite of 3 - 4 samples taken during steady state portion of run.

Analysis by Commercial Testing & Engineering (CT&E)

APPENDIX H. COAL SULFUR ANALYSIS

FluiDyne Run No.	EPA Test No.	Sulfur (wt.%)	
		CT&E	FluiDyne
25	1	---	3.29
26	2	3.23	3.20
27	3	---	3.38
28	3	---	3.29
29	4	3.27	3.25
35	5	3.21	3.31, 3.45
37	6	3.48	3.56
43	7	3.71	3.54, 3.65
44	8	---	3.55, 3.58
38	9	---	3.45, 3.53
40	11	3.67	3.77
41	12	---	3.42, 3.78
45	13	3.68	---
45	14	3.65	---
46	15	3.46	---
46	16	3.44	---
47	17	3.42	---
48	18	3.79	---
49	19	3.48	---
50	20	3.60	---
52	21	3.57	---
53	23	3.54, 3.62	---
54	24	3.74	---
55	25	3.56	---
56	26	3.75	---
57	27	3.75	---

CT&E - Commercial Testing & Engineering

APPENDIX I. BED AND FLY ASH SOLIDS ANALYSIS

FluiDyne Run No.	Source	Weight Percent Ca	 S	 SO_4	 CO_3	 Total C
25 (Test 1)	Bed	33.2	9.36	30.05	0.70	-
	Fly Ash	20.9	3.22	9.02	1.15	11.54
26 (Test 2)	Bed	34.3	9.66	32.57	0.57	-
	Fly Ash	19.44	3.62	10.55	0.35	8.89
28 (Test 3)	Bed	34.59	12.23	36.61	0.71	-
	Fly Ash	16.29	3.62	10.62	0.89	9.32
29 (Test 4)	Bed	32.88	11.7	34.97	0.64	-
	Fly Ash	22.01	5.44	16.27	2.09	10.23

APPENDIX I. BED AND FLY ASH SOLIDS ANALYSIS (CONT'D)

FluiDyne Run No.	Source	Ca(1)	Ca(2)	S(1)	S(2)	SO_4(1)	CO_3(1)	TOTAL C(1)
		Weight Percent						
35 (Test 5)	Bed	36.9	36.0	11.6	11.9	34.9	0.8	–
	Fly Ash	19.2	20.6	4.2	4.9	12.7	1.4	9.6
37 (Test 6)	Bed	35.2	29.1	10.0	8.0	29.7	3.1	–
	Fly Ash	19.9	–	5.2	–	15.2	1.5	7.5
38 (Test 9)	Bed	39.7	37.8	9.2	9.2	27.0	0.6	–
	Fly Ash	20.1	19.4	3.2	3.6	9.6	0.6	9.5
40 (Test 11)	Bed	39.2	38.0	10.4	9.1	31.1	0.7	–
	Fly Ash	18.0	16.7	3.7	4.2	10.7	0.5	8.8
41 (Test 12)	Bed	40.2	41.0	7.6	6.9	22.2	5.9	–
	Fly Ash	31.6	30.7	4.0	4.6	10.1	0.7	6.0
43 (Test 7)	Bed	39.2	35.9	11.1	10.2	32.4	1.0	–
	Fly Ash	30.3	30.0	7.4	8.5	22.0	0.6	7.6
44 (Test 8)	Bed	37.5	35.8	12.3	10.5	36.1	0.4	–
	Fly Ash	27.1	26.6	7.3	7.9	21.7	0.7	8.1

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(1) Analysis by CT&E (Commercial Testing & Engineering)

(2) Analysis by FluiDyne Engineering Corporation

APPENDIX I. BED AND FLY ASH SOLIDS ANALYSIS (CONT'D)

FluiDyne Run No.	Source	Weight Percent Ca	S	SO$_4$	CO$_3$	TOTAL C
45 (Test 13)	Bed	35.5	12.7	37.8	1.2	–
	Fly Ash	23.6	4.7	13.8	3.5	12.2
45 (Test 14)	Bed	37.5	11.6	34.4	1.2	–
	Fly Ash	22.8	4.3	12.6	2.6	13.0
46 (Test 15)	Bed	–	–	–	–	–
	Fly Ash	31.9	7.3	21.9	1.3	7.9
46 (Test 16)	Bed	36.4	11.7	34.5	0.9	–
	Fly Ash	31.9	6.7	19.9	1.2	6.4
47 (Test 17)	Bed	36.3	12.1	36.2	1.4	–
	Fly Ash	26.0	6.7	20.0	1.6	9.0
48 (Test 18)	Bed	–	–	–	–	–
	Fly Ash	24.9	6.7	19.5	0.8	10.5
49 (Test 19)	Bed	37.5	11.5	34.2	0.9	–
	Fly Ash	28.6	6.7	19.7	0.2	9.1
50 (Test 20)	Bed	36.8	11.8	35.0	1.1	–
	Fly Ash	20.0	3.4	10.0	1.7	13.3

APPENDIX I. BED AND FLY ASH SOLIDS ANALYSIS (CONT'D)

		Weight Percent				
Fluidyne Run No.	Source	Ca	S	SO_4	CO_3	TOTAL C
52 (Test 21)	Bed	34.0	11.2	32.1	2.4	-
	Fly Ash	14.0	3.6	10.3	2.4	10.5
53 (Test 22)	Bed	-	-	-	-	-
	Fly Ash	23.2	4.0	10.7	1.2	8.8
53 (Test 23)	Bed	40.3	9.3	27.5	3.2	-
	Fly Ash	22.9	4.0	11.9	1.2	5.6
54 (Test 24)	Bed	34.6	12.2	35.4	0.6	-
	Fly Ash	24.9	6.9	19.0	0.9	5.9
55 (Test 25)	Bed	35.1	13.0	38.6	0.4	-
	Fly Ash	22.6	6.7	19.9	2.3	6.0
56 (Test 26)	Bed	35.3	11.5	34.0	2.8	-
	Fly Ash	18.5	3.9	11.6	1.2	8.6
57 (Test 27)	Bed	37.9	11.9	34.6	3.0	-
	Fly Ash	20.7	4.5	11.5	1.4	7.1
57 (Test 28)	Bed	36.5	11.7	33.8	3.6	-
	Fly Ash	21.4	4.4	13.0	3.4	7.9

APPENDIX J. BED AND FLY ASH COMPOSITION AND COLLECTION RATE

Run No.	Source	Collection Rate, kg/hr	Composition (Wt %)			Calcium Fractions (Mole %)		
			Combustible Carbon	Ash	Sorbent	CaO	$CaCO_3$	$CaSO_4$
Run 25	Bed	44.0 (Meas.)[1] 49.7 (Calc.)	-	8.6	91.4	60.9	1.4	37.7
(Test 1)	Fly ash	57.2	11.3	46.1	42.6	78.3	3.7	18.0
Run 26	Bed	57.2 (Meas.) 46.3 (Calc.)	-	6.4	93.6	59.3	1.1	39.6
(Test 2)	Fly ash	49.0	8.8	49.0	42.2	76.2	1.2	22.6
Run 28	Bed	41.8 (Meas.) 35.7 (Calc.)	-	2.8	97.2	55.9	1.4	44.1
(Test 3)	Fly ash	52.2	9.1	54.4	36.5	71.4	1.1	29.0
Run 29	Bed	28.1 (Meas.) 27.6 (Calc.)	-	7.1	92.9	54.4	1.3	44.3
(Test 4)	Fly ash	60.3	9.8	35.8	54.4	62.9	6.3	30.8

(1) Calculated bed overflow rates are based on a calcium balance at a constant bed height.

APPENDIX J. BED AND FLY ASH COMPOSITION AND COLLECTION RATE (CONT'D)

Run No.	Source	Collection Rate, kg/hr	Composition (Wt %) Combustible Carbon	Ash	Sorbent	Calcium Fractions (Mole %) CaO	$CaCO_3$	$CaSO_4$
35 (Test 5)	Bed	21.2 (Meas.)(1) 21.1 (Calc.)	-	4.4	95.6	59.2	1.4	39.4
	Fly ash	41.4	9.6	44.9	45.5	67.5	4.9	27.6
37 (Test 6)	Bed	6.7 (Meas.) 20.5 (Calc.)	-	1.5	98.5	58.9	5.9	35.1
	Fly ash	35.8	7.5	43.8	48.7	63.2	5.0	31.8
38 (Test 9)	Bed	65.7 (Meas.) 66.5 (Calc.	-	0	103.4	70.7	1.0	28.3
	Fly ash	78.2	9.5	46.4	44.1	78.2	1.9	19.9
40 (Test 11)	Bed	83.1 (Meas.) 73.7 (Calc.)	-	0	106	65.7	1.2	33.1
	Fly ash	60.9	8.8	52.5	38.7	73.3	1.9	24.8
41 (Test 12)	Bed	100.5 (Meas.) 85.8 (Calc.)	-	0	104.3	67.2	9.8	23
	Fly ash	60.9	6.0	29.1	64.9	85.2	1.5	13.3
43 (Test 7)	Bed	0 (Meas.) 4.2 (Calc.)	-	0	107.0	63.9	1.7	34.4
	Fly ash	133.6	7.6	15.4	77	68.4	1.3	30.2
44 (Test 8)	Bed	-18.2 (Meas.) -17.9 (Calc.)	-	0	106.4	59.2	0.7	40.1
	Fly ash	139.5	8.1	21.1	70.7	64.9	1.7	33.4

APPENDIX J. BED AND FLY ASH COMPOSITION AND COLLECTION RATE (CONT'D)

Run No.	Source	Collection Rate, kg/hr	Composition (Wt %) Combustible Carbon	Composition (Wt %) Ash	Composition (Wt %) Sorbent	Calcium Fractions (Mole %) CaO	Calcium Fractions (Mole %) $CaCO_3$	Calcium Fractions (Mole %) $CaSO_4$
45 (Test 13)	Bed	47.8 (Meas.)(1) 34.4 (Calc.)	-	0	104	53.1	2.2	44.7
	Fly ash	89.8	12.2	29.0	58.8	65.7	9.9	24.4
45 (Test 14)	Bed	24.5 (Meas.) 89.5 (Calc.)	-	0	106	59.7	2.1	38.2
	Fly ash	62.5	13.0	33.4	53.6	69.4	7.6	23.0
46 (Test 15)	Bed	0 (Meas.) -12.5 (Calc.)	-	-	-	-	-	-
	Fly ash	149.6	7.9	10.1	82.0	68.7	2.7	28.6
46 (Test 16)	Bed	0 (Meas.) 0 (Calc.)	-	0	103	58.9	1.6	39.5
	Fly ash	107.1	6.4	13.6	80.0	71.5	2.5	26.0
47 (Test 17)	Bed	6.5 (Meas.) 5.9 (Calc.)	-	0	105	55.8	2.6	41.6
	Fly ash	81.4	9.0	23.3	67.7	63.9	4.1	32.0
48 (Test 18)	Bed	- (Meas.) 1.9 (Calc.)						
	Fly ash	128.2	10.5	24.4	65.0	65.3	2.1	32.6

APPENDIX J. BED AND FLY ASH COMPOSITION AND COLLECTION RATE (CONT'D)

			Composition (Wt %)			Calcium Fractions (Mole %)		
Run No.	Source	Collection Rate, kg/hr	Combustible Carbon	Ash	Sorbent	CaO	$CaCO_3$	$CaSO_4$
49 (Test 19)	Bed	0 (Meas.) -12.7 (Calc.)	-	0	105	60.4	1.6	38.0
	Fly ash	175.9	9.1	18.1	72.8	70.7	0.5	28.7
50 (Test 20)	Bed	54.6 (Meas.) 69.5 (Calc.)	-	0	105	58.4	2.0	39.6
	Fly ash	66.4	13.3	41.0	45.5	73.5	5.7	20.8
52 (Test 21)	Bed	- (Meas.) 31.6 (Calc.)	-	2.5	97.5	56.0	4.7	39.3
	Fly ash	35.5	10.5	55.0	34.5	57.9	11.4	30.7
53 (Test 22)	Bed	- (Meas.) 49.1 (Calc.)	-	-	-	-	-	-
	Fly ash	53.2	8.8	38.4	52.8	77.4	3.4	19.2
53 (Test 23)	Bed	- (Meas.) 27.0 (Calc.)	-	0	107	66.3	5.3	28.4
	Fly ash	48.6	5.6	40.6	53.8	74.9	3.5	21.6
54 (Test 24)	Bed	16.8 (Meas.) 12.6 (Calc.)	-	0	100	56.3	1.1	42.6
	Fly ash	62.5	5.9	29.7	64.4	65.9	2.4	31.7

APPENDIX J. BED AND FLY ASH COMPOSITION AND COLLECTION RATE (CONT'D)

Run No.	Source	Collection Rate, kg/hr	Composition (Wt %)			Calcium Fractions (Mole %)		
			Combustible Carbon	Ash	Sorbent	CaO	CaCO$_3$	CaSO$_4$
55 (Test 25)	Bed	-	-	0	104	53.4	0.8	45.8
	Fly ash	15.7 (Meas.) 43.0 (Calc.)	6.0	33.5	60.4	59.1	4.2	36.7
56 (Test 26)	Bed	-	-	0	104	56.6	3.3	40.1
	Fly ash	26.4 (Meas.) 30.0 (Calc.)	8.6	48.0	43.4	69.6	4.3	26.1
57 (Test 27)	Bed	-	-	0	108	56.7	5.3	38.0
	Fly ash	35.5 (Meas.) 35.1 (Calc.)	7.1	44.9	48.0	72.4	4.5	23.1
57 (Test 28)	Bed	-	-	0	105	54.8	6.6	38.6
	Fly ash	38.9 (Meas.) 36.7 (Calc.)	7.9	41.0	51.1	64.1	10.6	25.3

APPENDIX K. SULFUR BALANCES

FluiDyne Run Number	Baghouse Collection (kg/hr)	Bed Overflow (kg/hr)	Flue Gas (kg/hr)	Sulfur Found (kg/hr)	Sulfur in (kg/hr)	Balance (%)
25 (Test 1)	1.8	4.6	0.5	6.9	7.9	87
26 (Test 2)	1.8	4.5	0.6	6.9	7.6	91
28 (Test 3)	1.9	4.4	0.5	6.8	8.0	85
29 (Test 4)	3.3	3.2	0.6	7.1	8.2	87
35 (Test 5)	2.1	2.5	0.6	5.2	5.7	91
37 (Test 7)	2.0	2.0	0.6	4.6	5.7	81
38 (Test 9)	2.8	6.2	1.0	10.0	11.2	89
40 (Test 11)	2.5	7.7	0.7	10.9	12.3	89
41 (Test 12)	2.8	6.5	1.1	10.4	10.9	95
43 (Test 7)	11.3	0.5	0.7	12.5	11.9	105
44 (Test 8)	10.9	-2.2	1.4	10.1	11.2	90
45 (Test 13)	9.2	9.6	2.4	21.3	25.6	83
45 (Test 14)	5.9	23.0	2.7	31.6	28.5	111
46 (Test 15)	24.0	-3.0	0.7	21.7	25.3	86
46 (Test 16)	15.8	0	0.2	16.1	19.8	81
47 (Test 17)	12.0	1.6	1.3	14.9	19.5	76
48 (Test 18)	18.8	0.5	3.8	23.1	27.0	86
49 (Test 19)	25.9	-3.2	2.4	25.1	26.2	96
50 (Test 20)	5.0	18.0	2.3	25.3	25.2	100

APPENDIX K. SULFUR BALANCES (CONT'D)

FluiDyne Run Number	Baghouse Collection (kg/hr)	Bed Overflow (kg/hr)	Flue Gas (kg/hr)	Sulfur Found (kg/hr)	Sulfur in (kg/hr)	Balance (%)
52 (Test 21)	2.8	7.8	1.1	11.7	13.0	90
53 (Test 22)	4.7	13.2	1.2	19.2	18.7	103
53 (Test 23)	4.3	5.5	0.8	10.5	14.1	75
54 (Test 24)	9.5	3.4	0.5	13.4	14.3	94
55 (Test 25)	6.3	4.5	0.7	11.5	13.0	89
56 (Test 26)	2.6	6.7	1.3	10.6	12.8	83
57 (Test 27)	3.5	9.3	0.6	13.4	13.7	98

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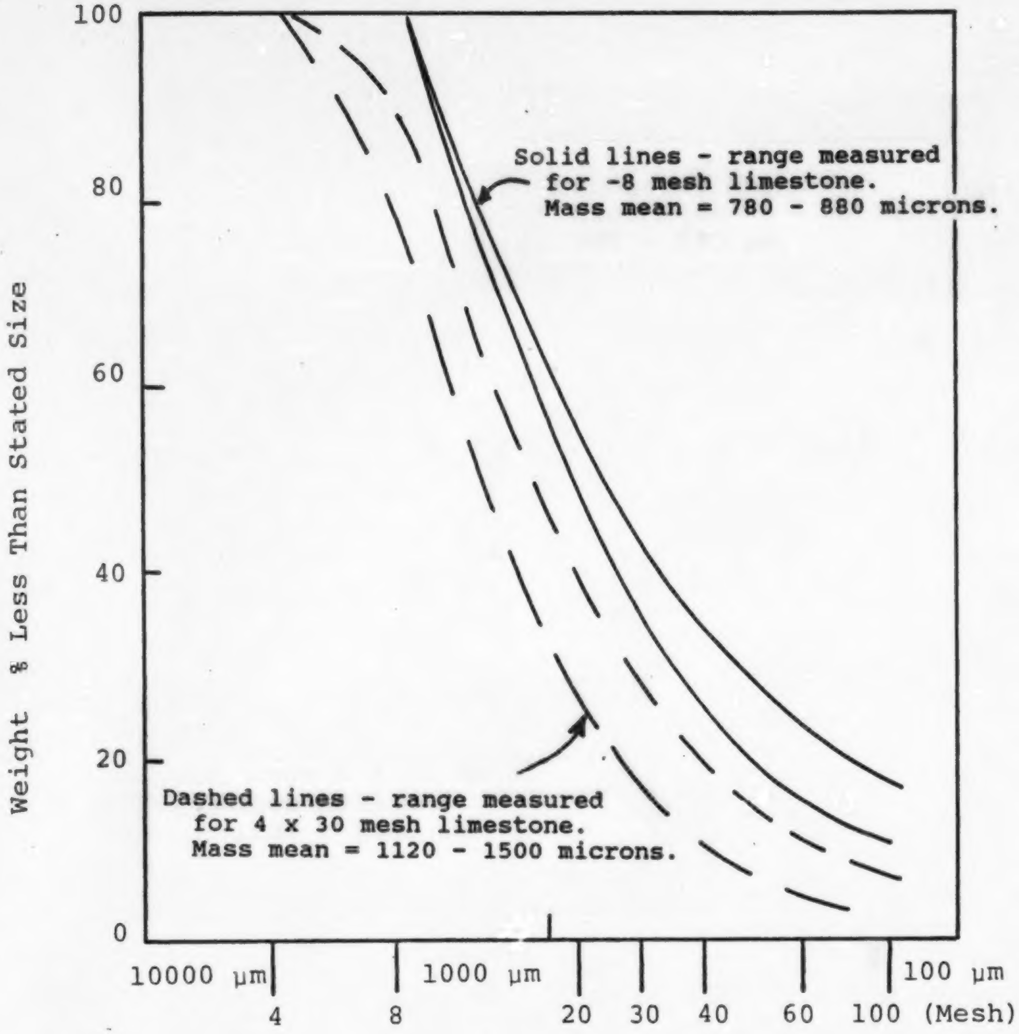
Average 90 ± 9

APPENDIX L. ILLINOIS COAL SIZE DISTRIBUTION

FluiDyne Run No.	EPA Test No.	Weight Percent Within the Stated Mesh Size						
		-3/4 - 8	8-20	20-30	30-40	40-60	60-100	<100
37	6	24	32	11	6	9	7	12
38	9	26	33	9	6	9	5	11
40	11	22	28	16	2	12	8	12
41	12	12	24	11	9	15	10	19
43	7	19	30	11	6	12	8	14
44	8	32	31	8	6	8	5	10
45	13	27	28	8	7	10	7	13
49	18	20	25	9	8	12	9	17
53	23	9	16	8	8	15	13	31
57	28	24	28	8	7	10	7	16

APPENDIX M. GREER LIMESTONE SIZE DISTRIBUTION

FluiDyne Run No.	EPA Test No.	Weight Percent Within the Stated Mesh Size						
		4-8	8-20	20-30	30-40	40-60	60-100	<100
26	2	0	47	8	10	10	8	17
35	5	0	49	12	8	10	5	16
38	9	21	54	10	6	4	1	3
40	11	21	52	10	6	5	1	5
41	12	7	49	17	9	8	3	7
43	7	8	51	15	10	8	2	6
44	8	15	45	16	9	8	2	5
45	14	19	48	12	7	7	3	5
51	21	0	62	14	8	7	3	6
54	24	0	49	14	10	10	6	11
57	28	0	48	11	8	11	7	15



APPENDIX M - GREER LIMESTONE SIZE DISTRIBUTION
(Graphical presentation)

APPENDIX N. BED OVERFLOW SIZE DISTRIBUTION

FluiDyne Run No.	EPA Test No.	Weight Percent Within the Stated Mesh Size						
		4-8	8-20	20-30	30-40	40-60	60-100	< 100
25	1	0	38	21	15	7	4	15
26	2	0	42	19	15	8	2	14
28	3	0	48	20	14	10	3	5
29	4	0	35	23	18	12	3	9
35	5	1	49	23	13	13	1	0
37	6	3	36	22	16	17	3	3
38	9	4	39	19	13	9	4	12
40	11	3	44	20	14	9	2	7
41	12	8	55	19	12	4	0	2
43	7	7	62	19	9	3	0	0
44	8	32	54	11	3	0	0	0
45	13	1	39	28	17	13	2	0
46	16	10	59	19	9	2	1	0
47	17	13	61	16	8	2	0	0
49	19	7	63	14	5	4	2	4
50	20	6	64	16	7	4	1	1
52	21	1	26	21	18	21	7	7
53	22	0	40	25	16	15	3	0
53	23	1	42	21	15	16	3	2
54	24	0	45	24	16	14	0	0

APPENDIX N. BED OVERFLOW SIZE DISTRIBUTION (CONT'D)

FluiDyne Run No.	EPA Test No.	Weight Percent Within the Stated Mesh Size						
		4-8	8-20	20-30	30-40	40-60	60-100	<100
55	25	1	53	21	14	10	0	0
56	26	1	39	22	15	15	4	4
57	27	2	35	26	18	16	2	2
57	28	2	43	23		15	1	2

APPENDIX O. FLY ASH SIZE DISTRIBUTION

FluiDyne Run No.	EPA Test No.	Weight Percent Within the Stated Mesh Size						
		>30	30-40	40-60	60-100	100-200	200-325	<325
25	1	10	6	10	10	10	11	43
26	2	4	2	4	6	7	7	70
28	3	0	2	1	3	5	7	82
29	4							
35	5	0	10	8	9	14	12	56
37	6	11	5	8	9	14	10	43
38	9	1	2	4	4	7	10	72
40	11	0	1	1	1	10	10	77
41	12	7	4	6	5	31	22	25
43	7	23	13	10	5	9	16	24
44	8	26	9	8	5	11	11	30
45	13	3	2	8	7	19	12	48
45	14	5	3	9	7	29	14	34
46	15	27	13	12	6	9	6	27
46	16	23	9	10	8	12	6	32
47	17	20	8	10	7	11	7	37
48	18	23	7	6	3	6	5	50
49	19	20	6	7	6	12	8	41

APPENDIX O. FLY ASH SIZE DISTRIBUTION (CONT'D)

FluiDyne Run No.	EPA Test No.	Weight Percent Within the Stated Mesh Size						
		>30	30-40	40-60	60-100	100-200	200-325	<325
50	20	0	0	2	3	14	11	70
52	21	0	0	1	8	20	11	60
53	22	0	1	6	10	21	10	52
53	23	0	1	3	9	20	10	57
54	24	8	6	16	11	14	7	38
56	26	0	1	2	11	22	12	53
57	28	2	1	8	10	17	9	53

APPENDIX P. FLUIDIZED BED BULK DENSITY

FluiDyne Run No.	EPA Test No.	Bulk Density Kg/dm^3
25	1	0.943
26	2	0.907
28	3	0.860
29	4	1.00
35	5	0.947
37	6	1.085
38	9	0.751
39	10	0.749
40	11	0.895
41	12	0.711
43	7	0.827
44	8	0.858
45	13	0.805
45	14	0.833
46	15	0.813
46	16	0.918
47	17	0.979
48	18	0.907
49	19	0.891
50	20	0.931
51 & 52	21	0.966
52	22	0.886
53	23	0.924
54	24	1.061
55	25	1.060
56	26	1.015
56 & 57	27	1.006
57	28	0.88